AFWAL-TR-81-2116

AD A114531

USAF SHALE OIL TO FUELS, PHASE II

UOP INC.
ALGONQUIN AND MT. PROSPECT ROADS
DES PLAINES, ILLINOIS 60016

NOVEMBER 1981

INTERIM REPORT FOR PERIOD 1 APRIL 1979 - 30 SEPTEMBER 1980

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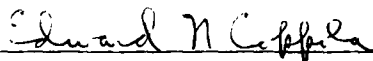
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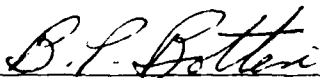
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REPORT DOCUMENTATION PAGE	READ INSTRUCTIONS BEFORE COMPLETING FORM

1. REPORT NUMBER AFWAL-TR-81-2116	2. GOVT ACCESSION NO. AO-A114 531	3. RECIPIENT'S CATALOG NUMBER

4. TITLE *(and Subtitle)* UNITED STATES AIR FORCE SHALE OIL TO FUELS - PHASE II INTERIM REPORT	5. TYPE OF REPORT & PERIOD COVERED INTERIM TECHNICAL REPORT 1 APR 79 - 30 SEP 80
	6. PERFORMING ORG. REPORT NUMBER

7. AUTHOR(*s*) J. R. Wilcox, J. G. Sikonia, T. G. Board, F. J. Riedl	8. CONTRACT OR GRANT NUMBER(*s*)

9. PERFORMING ORGANIZATION NAME AND ADDRESS UOP Process Division, UOP Inc. 20 UOP Plaza Des Plaines IL 60016	10. PROGRAM ELEMENT, PROJECT, TASK AREA & WORK UNIT NUMBERS Program Element 63215F Project 28400003

11. CONTROLLING OFFICE NAME AND ADDRESS Aero Propulsion Laboratory (AFWAL/POS) Air Force Wright Aeronautical Laboratories, AFSC Wright-Patterson Air Force Base, Ohio 45433	12. REPORT DATE November 1981
	13. NUMBER OF PAGES 295

14. MONITORING AGENCY NAME & ADDRESS*(if different from Controlling Office)*	15. SECURITY CLASS. *(of this report)* UNCLASSIFIED
	15a. DECLASSIFICATION/DOWNGRADING SCHEDULE

16. DISTRIBUTION STATEMENT *(of this Report)*

Approved for public release; distribution unlimited.

17. DISTRIBUTION STATEMENT *(of the abstract entered in Block 20, if different from Report)*

18. SUPPLEMENTARY NOTES

19. KEY WORDS *(Continue on reverse side if necessary and identify by block number)*

Shale Oil Refining	Denitrogenation
Hydrotreating	
Hydrocracking	
JP-4 Jet Fuel	
JP-8 Jet Fuel	

20. ABSTRACT *(Continue on reverse side if necessary and identify by block number)*

   Phase II of this project was conducted to demonstrate innovative technology to reduce the cost of converting shale oil to high yields of aviation turbine fuels. To carry out this program, UOP selected a processing scheme involving hydrocracking as the primary conversion unit. The Phase II program included pilot plant processing, fouling studies and economic analysis.

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The pilot plant operations performed during Phase II involved four specific processing steps: feed preparation, low pressure hydrotreating, high pressure hydrotreating, and hydrocracking. Two shale oil feedstocks were utilized for each processing step: shale oil derived from Occidental Modified In-Situ retort and shale oil obtained from the Paraho direct-heated retort.

Using data generated from the pilot plant operations and the study basis provided by the USAF, economic analyses were performed to find the cost of production of 1) jet fuel, and 2) total liquid product at a 15% discounted cash flow rate of return (DCFRR) on investment. Assuming 100,000 barrels per stream day (BPSD) of raw shale oil valued at $30/bbl, the following results were obtained:

	Max. JP-4	Max. JP-8
Jet Fuel Yield, BPSD	94,420	82,400
Total Liquid Product Yield, BPSD	94,420	90,520
Capital Investment, MM $	695	726
Product Costs at 15% DSFRR		
$/Bbl of Total Liquid Product	42.971	45.50

FOREWORD

The interim report describes the work completed by UOP Inc. on Phase II of the Contract No. F33615-78-C-2079 entitled "United States Air Force Shale Oil to Fuels."

The work reported herein was performed during the period 1 April 1979 - 30 September 1980 under the direction of A. O. Braun and J. R. Wilcox. This report was submitted to the USAF in November, 1981.

The following individuals contributed to the execution of the program and the preparation of the report material:

> A. O. Braun, F. H. Franke, Lee Hilfman, E. J. Latos and
> F. J. Riedl; all of the Corporate Research Center, UOP Inc.;
> T. G. Board, V. A. Gembicki, J. G. Sikonia, J. R. Wilcox and
> Edwin Yuh; all of the Process Division of UOP Inc.

A final report summarizing the Phase III and Phase IV tasks will follow.

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SUMMARY

Phase II of this project was conducted to demonstrate innovative technology to reduce the cost of converting shale oil to high yields of aviation turbine fuels. To carry out this program, UOP selected a processing scheme involving hydrocracking as the primary conversion unit. The Phase II program included pilot plant processing, fouling studies and economic analysis.

The pilot plant operations performed during Phase II involved four specific processing steps: feed preparation, low pressure hydrotreating, high pressure hydrotreating, and hydrocracking. Two shale oil feedstocks were utilized for each processing step: shale oil derived from Occidental Modified In-Situ retort and shale oil obtained from the Paraho direct-heated retort.

A total of 3700 gallons of raw Occidental and 400 gallons of raw Paraho shale oil was desalted and dewatered during the feed preparation step for use as feedstock to the low pressure hydrotreating operation.

A first-stage hydrotreating process variable and catalyst study was performed to determine the optimum operating conditions and catalyst system necessary to reduce the arsenic level to less than one ppm while saturating the diolefins and olefins. Using these conditions, approximately 2,200 gallons of raw desalted Occidental and 300 gallons of raw desalted Paraho shale oils were successfully processed.

Using the arsenic free shale oil feedstock, a process variable and catalyst study was again conducted to determine the optimum operating conditions and catalyst system required to reduce the total nitrogen level so that the shale oil could be utilized as hydrocracker feedstock. Based on the variable study, three 100 gallon batches of shale oil were produced with nitrogen contents varying from 100 to 900 ppm to serve as feed for the hydrocracking step.

The hydrocracking pilot plant operation established the advantages of a novel process flow arrangement and demonstrated the required commercial hydrocracking process conditions for the production of jet fuel meeting the JP-4 and JP-8 specifications. Samples of selected fuels representative of the products from the proposed flow arrangement were produced and supplied to the USAF.

The shale oil fouling study, which is to continue in Phase III, showed that the Monirex Fouling Monitor could be successfully used to evaluate the relative fouling rates of shale derived oils and petroleum derived fractions. Above 250°C, raw Occidental shale oil showed a higher fouling rate than Paraho.

Using data generated from the pilot plant operations and the study basis provided by the USAF, economic analyses were performed to find the cost of production of 1) jet fuel, and 2) total liquid product at a 15% discounted cash flow rate of return (DCFRR) on investment. Assuming 100,000 barrels per stream day (BPSD) of raw shale oil valued at $30/bbl, the following results were obtained:

	Max. JP-4	Max. JP-8
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Capital Investment, MM $	695	726
Product Costs at 15% DCFRR, $/Bbl of Total Liquid Product	42.97	45.50

SECTION I

INTRODUCTION

Background and Objectives

This study was conducted to develop yield and economic data for the conversion of shale oil to aviation turbine fuels using innovative processing techniques. A processing scheme maximizing the yield of jet fuels at minimum cost was developed during Phase I of this program. This scheme consisted of the primary shale oil upgrading and conversion units with the necessary auxiliary facilities. The primary units include two stages of upgrading utilizing proprietary RCD Unibon® technology combined with UOP Hydrotreating technology. The upgraded shale oil is then converted to high yields of jet fuel using a novel, "parallel flow" hydrocracking operation included in the UOP proprietary HC Unibon® technology.

Phase I of the program provided for the development of innovative processing techniques and associated information for the conversion of shale oil to high yields of jet fuel. Phase II was primarily a test program utilizing small scale pilot plants to demonstrate the processing techniques developed in Phase I. The following summarizes the specific objectives for Phase II:

- Characterization of feedstock

- Determination of the shale oil fouling characteristics

- Preparation of sufficient feedstock to explore "parallel flow" hydrocracking operating parameters

- Exploration of first and second stage hydrotreating operating parameters and catalyst systems

- Production of small scale jet fuel product samples

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- Verification and/or adjustment of the operating requirement and product quality estimates generated during Phase I

- Establishment of processing parameter information for the Phase IV Economic Evaluation Study

Two feedstocks were used in this phase of the project. The first was shale oil derived from Occidental Petroleum Company's Modified In-Situ retorting process, and the second was shale oil obtained from the above-ground Paraho direct-heated retort.

The pilot plant operations performed during Phase II involved the following specific processing steps: feed preparation (desalting), two upgrading steps (two stages of hydrotreatment), and conversion to fuels (hydrocracking).

As shown in Figure 1, the UOP approach to the problem of shale oil conversion to high quality fuels involves three distinct hydroprocessing steps (after the raw shale oil has been properly desalted). Shale oil has unique characteristics relative to conventional crude petroleum oil. Unusually high arsenic and iron levels, high pour point and viscosity, a high unsaturates concentration, and a significant solids (ash) content make conventional front-end refining techniques unusable without proper pretreatment.

The initial step in the UOP scheme is a feed preparation procedure which is essentially a desalting operation. The pilot plant utilized for this operation is completely analogous to a two-stage desalting unit commonly used in the refining industry.

After desalting, the first pretreatment step involves the use of UOP's RCD Unibon technology to effect metals removal, sulfur reduction, a degree of diolefin and olefin saturation, and the final solids clean-up necessary to render the resulting effluent suitable for subsequent processing.

Another characteristic of shale oil is its high nitrogen content relative to conventional crude oil. Reducing this contaminant to acceptable levels requires the use of a second stage, high severity hydrotreatment of the first-stage hydrotreater effluent.

Once the metals and nitrogen contents have been reduced to acceptable levels and the unsaturates hydrogenated, shale oil is ready for the primary conversion step -- hydrocracking to jet and other fuels. The hydrocracking process that UOP has selected is a parallel flow hydrocracker developed for conventional petroleum use. One of the objectives of Phase II was to determine whether the advantages of this particular configuration, evident in petroleum processing, are achievable in shale oil processing.

The overall refinery block flow diagram of the UOP approach used in the economic evaluations is shown in Figure 2. Although jet fuel is the primary product, diesel fuel and naphtha were also produced. The naphtha was split into light and heavy fractions, with the heavy fraction converted to a gasoline component by reforming via a UOP Platforming process unit. The reformate was mixed with the light naphtha fraction and sufficient butanes to produce the finished gasoline.

In Figure 3 a more detailed block flow diagram is presented, this one specifically designed for production of JP-8 jet fuel. Hydrogen production is supplied from a steam reforming plant operating on light hydrocarbons. In addition to the fuels produced, sulfur and ammonia are also products of this operation.

In Figure 4 a block flow diagram for the production of JP-4 jet fuel is shown. The only significant difference between this scheme and that for JP-8 production is the elimination of the naphtha hydrotreating and Platforming units. When maximum JP-4 is desired, the heavy naphtha fraction, otherwise available for gasoline production, is fully utilized in the production of JP-4. When less than maximum conversion to JP-4 is desired, hydrotreating and reforming can be added to produce a gasoline product.

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The various operating conditions used during the pilot plant studies are consistent with commercial operating experience based on petroleum derived oils. All of the data presented in this report will have operating conditions reported relative to a base operating condition. In all cases, the base conditions selected are those required to process a typical Middle East petroleum of the same boiling range to the identical product requirements. Positive differences in operating conditions indicate more severe requirements; negative differences indicate less severe processing requirements.

PROBLEM	PROCESS
METALS REMOVAL (As, Fe)	
VISCOSITY REDUCTION	FIRST STAGE
STABILIZATION-DIOLEFIN SATURATION	HYDROTREATING
NITROGEN REMOVAL	SECOND STAGE
COMPLETE DESULFURIZATION	HYDROTREATING
CONVERSION TO JET, TRANSPORTATION,	
AND HEATING FUELS	HYDROCRACKING

UOP 663-45

FIGURE 1
SHALE OIL TO FUELS
UOP APPROACH
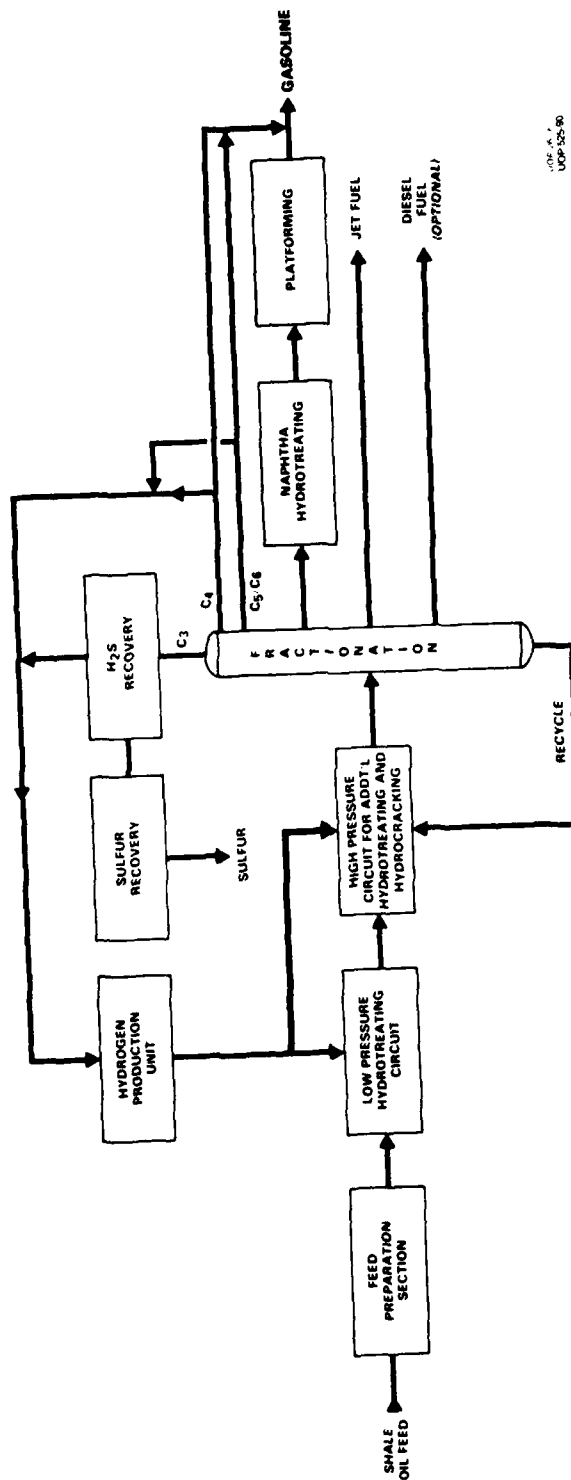
FIGURE 2

UOP SHALE OIL TO FUELS

OVERALL BLOCK FLOW DIAGRAM

SULFUR

SULFUR PLANT

H_2S FROM SOUR WATER TREATER

H_2S

HYDROGEN TO PROCESS UNITS

STEAM REFORMING

CLEAN FUEL GAS

AMINE TREATING

SOUR GAS FROM PROCESS UNITS

WHOLE SHALE OIL

FEED PREPARATION SECTION

RCD UNIBON FIRST STAGE HYDROTREATING

HYDROGEN

UOP HYDROTREATING SECOND STAGE HYDROTREATING

HYDROGEN

HC UNIBON (HYDROCRACKING)

HYDROGEN

SOUR GAS TO TREATING

RECYCLE

FRACTIONATION SECTION

SOUR GAS TO TREATING

SOUR BUTANES

HEAVY NAPHTHA

LIGHT NAPHTHA

TO HYDROGEN PRODUCTION

HYDROTREATER PLATFORMER

HYDROGEN RICH GAS

LIGHT NAPHTHA

BUTANES

GASOLINE

JET FUEL JP-8

DIESEL

FUEL OIL TO PROCESS

H_2S TO SULFUR PLANT

NH_3

SOUR WATER TREATMENT

AMMONIA

CLEAN WATER TO REFINERY

REFINERY FUEL OIL

FIGURE 3

PRODUCTION OF JP-8 JET FUEL
BLOCK FLOW DIAGRAM

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FIGURE 4

PRODUCTION OF JP-4 JET FUEL
BLOCK FLOW DIAGRAM

SECTION II

PILOT PLANT PROGRAM

This section describes the results of the various pilot plant operations outlined in the previous section. A description of the pilot plant equipment utilized during this program is also provided.

Feedstock Preparation

This initial pilot plant operation prepared the raw shale oil feed for downstream catalytic processing by removing essentially all of the water and reducing the solids in a conventional "desalting" operation. A total of 3700 gallons of raw Occidental and 400 gallons of raw Paraho shale oil was desalted during this operation. Detailed analytical inspections of both of the raw shale oils are shown in Tables 1 and 2. A schematic drawing of the pilot plant utilized for this processing step is shown in Figure 5.

The desalting operation of the raw Occidental shale oil was performed in two separate operations. During the initial operation, 1,250 gallons of the raw shale oil was processed in a once-through operation utilizing only the first-stage electrode. The operating specifications were based upon the manufacturer's recommended voltage drop across the electrode. This recommended voltage coincides with the condition of optimum water-oil phase separation. Temperature, pressure and water injection into the plant were varied until this voltage was achieved. The operating conditions used during this operation are shown in Table 3. Immediately following this once-through desalting operation, the plant was placed on total recycle allowing adequate time for sufficient water and sediment settling. Desalted shale oil was then decanted and transferred to 55 gallon drums

for storage. An analytical inspection of the raw and desalted shale oil from this initial operation is shown in Table 5.

The remaining 2,450 gallons of raw Occidental shale oil were treated in the same plant using both stages. Operating conditions for both stages are shown in Table 4. The plant was operated in the same manner as during the initial run. Analytical inspections of the desalted shale oil from this run are shown in Table 5. This material was then blended with the 1,250 gallons of desalted oil from the first run to provide the feedstock for the first-stage hydrotreating operation. An inspection of this final blend is also included in Table 5.

Raw Paraho shale oil (400 gallons) was treated in exactly the same manner as the Occidental material. Both stages of the plant were utilized as in the second Occidental run. Operating conditions and analytical inspections of the raw and desalted shale oil are shown in Tables 6 and 7, respectively.

As indicated in the analyses in Tables 5 and 7, this conventional desalting operation did appear to reduce the level of some contaminants, but was not highly efficient. For example, BS and W, heptane insolubles, and ash were moderately reduced, but not substantially eliminated. If complete removal of a contaminant such as solids (ash) is required, some additional processing steps would be necessary. Essentially no differences were observed between single and double stage desalting.

The variations noted in some analyses result from both normal differences in test results and special analytical problems, as in the determination of iron and oxygen levels.

First-Stage Hydrotreating

The desalted shale oil must be treated in a first treatment step or first-stage, low pressure hydrotreating unit to provide for the removal of certain contaminants and stabilization of the raw shale oil, to allow high severity hydrotreating and subsequent conversion operations.

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The first-stage hydrotreating pilot plant program involved three separate pilot plant phases. The objectives of each of these phases were:

1. Perform a detailed process variable study that included examining hydrotreating operating variables and various catalyst systems.

2. Confirm the operating conditions necessary to reduce the arsenic content of the shale oil to less than 1 ppm in a large pilot plant.

3. Hydrotreat approximately 2,200 gallons of raw desalted Occidental and approximately 300 gallons of desalted Paraho shale oil while examining the operating conditions in (2) above.

Process Variable Study

The first part of this hydrotreating study was conducted to examine the effects of operating variables and the performance of various catalysts. This study was designed:

1. To establish conditions for arsenic removal to a level of 1 ppm maximum in the product.

2. To determine the degree of olefin saturation achievable.

3. To select the best catalyst and conditions for first-stage hydrotreating.

Answers to the above points would provide more definitive information on shale oil processing for selection of conditions for the Phase III six-month catalyst stability test, as well as conditions for the preparation of feed for subsequent Phase III hydrocracking studies.

Conditions -- Based on the guidelines reviewed in the introduction, the process conditions used in the first-stage hydrotreating study are expressed relative to the conditions required to stabilize a corresponding Middle East petroleum fraction that contains a significant quantity of unsaturated components.

		Written as
Temperature	T-T (base), °F	$T-T_B$
Pressure	P-P (base), psi	$P-P_B$
Space Velocity	LHSV/LHSV (base)	L/L_B
Combined Feed Ratio	CFR/CFR (base)	C/C_B
Hydrogen Circulation	SCFB/SCFB (base)	H/H_B

The base conditions selected for reference represent those employed commercially for hydrotreating a typical Middle East petroleum gas oil.

The following range of conditions was investigated to provide answers to the program outlined above:

Temperatures	-400°F to +100°F
Pressures	0 to 450 psi
LHSV	0.16 to 0.5
Hydrogen Circulation	0.67 to 1.33 (once-through hydrogen)

Catalysts -- Commercial UOP catalysts employed in this study include the first-stage hydrotreating catalysts designated DSA and DRA. The use of the DSA catalyst was based on its application to olefinic petroleum stocks. The DRA catalyst was selected for its excellent hydrotreating ability in the presence of C_7 insolubles and its high metals removal capability. A third material, α-alumina, was used to check the extent of non-catalytic metal removal from the shale oil.

Equipment and Procedures -- A UOP Research pilot plant, specifically suited to this type of study, is shown schematically in Figure 6. It is a once-through unit and consists of a trickle bed reactor, a reactor heater, and a two-phase separator (flash drum). Feed facilities consist of weighed liquid and measured hydrogen systems.

Hydrogen and shale oil feed are mixed and pumped together downflow over the catalyst. The resulting product effluent is separated in the flash drum. The product off-gas is measured and the liquid product is weighed and analyzed. Sufficient line-out time was provided after every condition change before sampling the product. Test periods throughout the work were 8 hours in duration.

The following analytical tests were used in the work for control purposes: Bromine Number, API Gravity, Arsenic (ppm), and Iron (ppm).

Specific tests were selected for complete work-up and analysis.

<u>Charge Stocks</u> -- Analytical inspections of the desalted shale oils received at the UOP Research facilities for this study are listed in Tables 5 and 7.

Differences between the two shale oils are noted in several areas. The hydrogen content of Paraho shale oil is lower; the oxygen and nitrogen content, Conradson carbon and Bromine Number are higher than that in the Occidental shale oil. On the other hand, the shale oil produced from the Modified In-Situ Occidental process showed a higher ash and metal-contamination. It is not known if this would be typical of commercially produced materials.

<u>Process Studies</u> -- The Occidental and Paraho shale oils were processed over DSA catalyst in two separate runs. The Occidental shale was also processed over DRA catalyst in another run. In these operations, the effects of varying space velocity, pressure, temperature and hydrogen circulation were studied. The results are presented in Tables 8 through 9 and in Figures 7 through 11.

The largest response to olefin saturation was noted with temperature changes. Little saturation was observed at temperature lower than about 20°F below base average bed temperature (ABT), but temperatures above that level produced a sharp drop in the Bromine Number (Figures 7 and 8). Essentially little or no change in saturation was observed with either the

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DSA or DRA catalyst when increasing the hydrogen ratio from 0.67 to 1.33 times base hydrogen recycle ratio (Figures 7, 8 and 11). Changing space velocity from 0.16 to 0.5 times base and pressure from base to 450 psi above base produced only small changes in the saturation of either feed (Figures 9 and 10).

Arsenic and iron removal from the Occidental and Paraho shale oils closely resembles the olefin saturation picture. The data are presented in Tables 8 through 18 and summarized in Figures 12 through 17.

In terms of arsenic and iron removal, the process variable study for Occidental shale oil over DSA catalyst showed gradual reduction of iron and arsenic is rapid at temperatures higher than 300°F below base. The 1 ppm arsenic level was obtained at 20°F above base and 0.33 L/L_B (Table 9). At the operating conditions investigated, an increase from 0.67 to 1.33 SCFB times base hydrogen recycle ratio does not show any definitive change. Figures 14 and 15 show that the removal of sulfur and iron is directly related to arsenic removal. At high arsenic removal levels, the loss of resolution in the analytical methods makes the correlation of iron removal meaningless. Even though the data are somewhat limited by these analytical deficiencies, a kinetic study showed that the arsenic removal reaction is reasonably represented by first order kinetics (Figure 14). On this basis, and as shown in Figure 15, an activation energy (E_a) of 11,500 cal/mol was calculated. A similar analysis for iron removal showed an activation energy of 6,700 cal/mol.

Overall, the arsenic and iron removal from Paraho shale oil required less severe operating conditions to achieve a given product specification, however, this feed contained lower quantities of these contaminants. Based on the results using the DSA catalyst the conditions required to obtain a 1 ppm arsenic level in the product are:

A.   Occidental shale oil:

     0 psi, 1.33 SCFB hydrogen, +50°F ABT, 0.33 L/L$_R$.

B.   Paraho shale oil:

     0 psi, 1.33 SCFB hydrogen, +50°F ABT, 1.5 L/L$_R$.

When using the DRA catalyt, the conditions required to obtain 1 ppm iron and arsenic levels in the Occidental shale oil product are: 0 psi, 1.33 SCFB hydrogen, 0.33 L/L$_R$ and +50°F ABT (Table 18).

Inspections of the preheater α-alumina support material in Run 1 (Table 20) indicated iron and arsenic deposition. The use of such material for upstream metals removal would extend the useful life of the hydrotreating catalyst and may provide information as to the original state of the iron and arsenic. To confirm or refute the ability of a support material to remove arsenic and other metal contaminants, a separate run was carried out (Run 3) using only the α-alumina. The results of this operation are summarized in Table 14 while Figure 16 is a plot of the arsenic product content along with the similar data of Run 2 for comparison. These results show that a 40% reduction in arsenic and a 57% reduction in iron were achieved with the use of this material, but the much lower removal rates and lack of temperature response show that, as expected, the α-alumina is non-catalytic.

Some of the products in this non-catalytic run were analyzed and these data are shown in Table 15. Some reduction of sulfur occurs, approximately 17%, but no nitrogen removal is observed with the use of α-alumina. A small amount of non-catalytic hydrogenation of the product occurs as seen by the reduction in Bromine Number.

Analyses of the used catalyst and preheater material from all of the Runs, 1 through 5, are presented in Tables 20 through 24. Analyses include iron, arsenic, sulfur, carbon and nitrogen. The preheater material and catalyst were broken into sections and the above analyses

obtained on each section. The same pattern of higher to lower deposition at the preheater or catalyst inlet to outlet prevailed in all of the runs.

Catalyst Stability -- In the longest two operations, Runs 1 and 4, test data show good short term catalyst stability with regard to arsenic and iron removal using both the Occidental and Paraho feeds. Data abstracted from Tables 12, 16 and 17 are shown below.

Run 1 Catalyst DSA Occidental Shale Oil Feed
Conditions: 1.0 $P-P_B$, 0.67 H/H_B, 0.5 L/L_R

Hrs. On Stream	ABT Catalyst, °F	Arsenic, ppm	Iron, ppm	Bromine Number	Sulfur, %
Feed		27	42	23.6	0.64
136-144	-30	6	5.6	15.8	0.26
808-816	-30	6	2.2	15.7	0.23
840-848	-30	2	--	15.9	0.29

Run 4 Catalyst DSA Paraho Shale Oil Feed
Conditions: 1.0 $P-P_B$, 0.67 H/H_B, 0.5 L/L_R

Hrs. On Stream	ABT Catalyst, °F	Arsenic, ppm	Iron, ppm	Bromine Number
Feed		19	38	34.7
50-98	-25	2	2.8	23.8
442-506	-25	1	2.7	

No deactivation of the DSA catalyst was evident due to arsenic and iron deposition or poisoning. Product Bromine Number and sulfur content remained unchanged during the entire run. In addition, no plugging occurred in the reactor preheaters or catalyst zones during this processing period.

Selection of Catalyst and Conditions -- The initial processing study program called for the testing of at least two first-stage catalysts. In order to maintain the working schedule and because of its known high hydrogenation activity and contaminant removal ability, the DSA catalyst was the first to be tested.

The selection of conditions commensurate with the catalyst choice was as listed in the previous section, when processing Occidental feed to produce a 1 ppm arsenic content product:

$$0 \text{ P-P}_B, 0.33 \text{ L/L}_B, +50°F \text{ ABT}, 1.33 \text{ H/H}_B$$

A special run, No. 2, was made using the DSA catalyst to confirm the selection of conditions. A run summary is given in Table 13 and a plot of the arsenic product content and catalyst average bed temperature against hours on stream is shown in Figure 17.

The operation did confirm the conditions selected as noted above. Analytical or sampling problems are thought to be responsible for the two high points reported during the latter part of this run. Information indicated that pilot plant operating difficulties were experienced at the time of analysis.

The product from 230-458 hours was blended and sent for analyses. These analyses are shown in Table 19. These data also show the critical arsenic contaminant to be at 1 ppm level and a 65% reduction of olefinicity. The extremely low C_7 insolubles level indicates asphaltic content was nil. The sulfur, iron and nitrogen removals were approximately 92%, 94% and 43%, respectively.

The DRA catalyst subsequently tested in Run 5 (Table 18) was selected for the Phase III six-month stability test. This will provide additional information on the catalyst's ability to function long term with a metal containing feed for which it has been designed.

Product data were abstracted from Tables 9 and 18 for the DSA and DRA catalysts, respectively, and compared below at two space velocities.

Conditions: 0 P-P$_B$, 1.33 H/H$_B$

At 0.5 L/L$_R$ and -40°F ABT

Catalyst	API	Br. No.	Con. C., Wt-%	S, Wt-%	N, Wt-%	O$_2$, Wt-%	C/H, Wt-%	Fe/As, ppm
UOP-DSA	27.1	10.7	0.43	0.10	1.04	0.42	12.55/85.09	5.3/3.0
UOP-DRA	26.3	10.8	0.45	0.09	1.18	0.25	12.56/85.93	1.8/1.0

At 0.33 L/L$_R$ and +40°F ABT

Catalyst	API	Br. No.	Con. C., Wt-%	S, Wt-%	N, Wt-%	O$_2$, Wt-%	C/H, Wt-%	Fe/As, ppm
UOP-DSA	30.0	9.0	0.38	0.09	1.06	0.21	12.64/85.72	2.3/1.0
UOP-DRA	27.0	8.8	0.41	0.05	1.11	0.31	12.56/86.75	1.5/1.0

Based on the results, the performance of the DSA and DRA catalysts are comparable with some small differences. Future Phase III work should determine whether the DRA catalyst can tolerate solids and metals and still continue to function in hydrotreating.

Large Scale Production Run

The objective of this work was to produce substantial quantities of hydrotreated Occidental and Paraho shale oils for further process studies.

Catalyst -- Based on the data available from the variable study, the commercially available hydrotreating catalyst, DSA catalyst, was used to remove arsenic from both the desalted Occidental shale oil and the desalted Paraho shale oil. This catalyst is a commercially proven, high activity, Co-Mo extrudate.

Charge Stock -- The charge stock used during the first portion of the run was desalted Occidental shale oil. Table 5 shows the charge stock inspection of the composited Occidental shale oil used during the production run. The charge stock used during the second portion of the production run was desalted Paraho shale oil. The charge stock inspection of this material was shown in Table 7.

Equipment and Operations -- The pilot plant flow scheme used for the production run is presented in Figure 18. Fresh feed and recycle hydrogen were combined at the top of the reactor and flow concurrently downward through the catalyst bed. Water was injected at the reactor outlet to prevent the formation of solid salts. The mixture then flowed to the high pressure separator (HPS) where the gas and liquid phases were separated. The gas leaving the HPS went to an overhead condenser where additional water was injected. This mixture was then directed to a hot flash separator (HFS) where the water and gas were separated. The gas leaving the HFS was combined with make-up hydrogen and recycled back to the top of the reactor. The water leaving the HFS was sent to the caustic scrubbing system.

The water-oil mixture leaving the HFS flowed to a low pressure separator (LPS) where the oil and water were separated. The oil leaving the LPS went to a stripper column where any carry-over water or H_2S and NH_3 produced in the reactor were separated from the hydrotreated product. The water and gas leaving the stripper column and the water from the LPS were then sent to the caustic stripping system. The hydrotreated shale oil from the bottom of the stripper column was collected in inert gas blanketed drums.

Several modifications were made to the plant prior to the start of these runs. The cooling coil located at the outlet of the reactor was replaced with a straight section of steam-traced pipe. This allowed water to be injected at the reactor outlet. In addition, the gas line running from the LPS to the overhead condenser was steam traced and steam, rather than cooling water, used in this condenser.

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Due to the high arsenic content of shale oils, the formation of a significant concentration of arsine gas during the hydrotreating operation is possible. Arsine, AsH_3, is a highly poisonous, colorless, non-irritating gas. It may be formed when inorganic arsenic is contacted with nascent (freshly formed) hydrogen.

A bubbler system for the detection and removal of any arsine formed during the hydrotreating of the shale oil was also constructed. The caustic scrubbing system was repaired and modified so that any off-gas from the LPS, stripper column, product drums, spent caustic drums, and sample ports would be sent to the caustic scrubbing system.

Following the catalyst prewetting and sulfiding operations with light cycle oil, the reactor was then prewet with the charge stock and maintained liquid full for approximately five weeks while the bubbler system modifications mentioned above were completed. In the start-up period, the plant was lined-out on total recycle at the following initial operating conditions:

Plant Pressure, psi $(P-P_B)$	0
Fresh Feed LHSV (L/L_B)	0.15
H_2 Recycle Rate (H/H_B)	1.47
Reactor Temp., °F $(T-T_B)$ Ave.	+65

At these conditions, the bottom product of the stripper column was collected in drums which were changed when they contained approximately 45 gallons of product.

Analyses of the product collected during the first four weight checks of Run 1 showed arsenic levels substantially below 1.0 ppm. For this reason, and as shown in Figures 19 and 20, the charge rate was approximately doubled which brought the product arsenic content up to the 1 ppm level. The plant operating conditions for the remainder of Run 1 were as follows:

Reactor Pressure, psi $(P-P_B)$	0
Fresh Feed LHSV (L/L_B)	0.32
H_2 Recycle Rate (H/H_B)	1.33
Reactor Temp., °F $(T-T_B)$ Ave.	+65

At the end of the production run on Occidental shale oil, all the drums of hydrotreated product, except the line-out drum, were combined. A complete analytical inspection of this material was then taken. The results of this inspection are presented in Tables 25 and 26.

Immediately after Production Run 1 and without shutting the plant down, the desalted Paraho shale oil was charged to the plant. The hydrotreating of this material was designated Production Run 2 and is shown graphically in Figure 21. The first drum of hydrotreated product was labeled "Line-Out Material". The product was then collected in numbered drums corresponding to the particular weight check number. Because of the relatively small amount of material to be hydrotreated, the plant operating conditions for Production Run 2 were substantially the same as for the Occidental feedstock:

Plant Pressure, psi $(P-P_B)$	0
Fresh Feed LHSV (L/L_B)	0.31
H_2 Recycle Rate (H/H_B)	1.33
Reactor Temp., °F $(T-T_B)$ Ave.	+73

At the end of this run all of the drums of hydrotreated Paraho shale oil, except the line-out drum, were combined. An inspection of this material was then taken. The results of this inspection are presented in Tables 27 and 28.

A comparison of the two first-stage products shows that both met the requirement of containing 1 ppm or less of arsenic. As indicated in Tables 25 through 28 and as shown in Figures 20 and 21, there remains a discrepancy in the iron contents. The Occidental material showed a consistent, but unexpectedly high, iron content of about 13 ppm. A value of

about 3 ppm was expected from the previous work (Table 19). In both materials, the Bromine Number and viscosities were reduced by about 50%.

First-stage hydrotreating unit weight balances are given in Tables 29 and 30. About 580 and 765 SCFB of hydrogen were consumed in the Occidental and Paraho arsenic removal and stabilization operations, respectively.

An arsenic balance was performed for each of the above runs. From these calculations it can be seen that approximately 90 percent of the incoming arsenic can be accounted for either on the catalyst, in the product, or in the spent caustic.

After the desalted Paraho shale oil was processed in Production Run No. 2, the catalyst bed was flushed with light cycle oil (LCO) and unloaded into five equally sized beds that were later analyzed for iron, arsenic, carbon, nitrogen, sulfur and hydrogen content. The results of these tests are shown in Table 31. Both the iron and aresenic are concentrated in the top part of the catalyst bed. However, significant amounts of these elements are found throughout the bed.

No significant plugging of the external or internal reactor preheaters occurred during either run. In addition, no arsine was detected in the plant area, the hydrotreated product, or during unloading of the catalyst bed.

Summary -- Approximately 2200 gallons of desalted Occidental shale oil and 300 gallons of desalted Paraho shale oil were hydrotreated in a large pilot plant. Analyses of both the hydrotreated products also showed that along with achieving the desired arsenic level of 1 ppm or less, substantial reductions in lead, sulfur, nitrogen, ASTM ash, Bromine Number and C_7 insolubles also occurred during the hydrotreating process. The arsenic specification was met at both 0.15 and 0.32 L/L_B during Production Run 1.

No significant plugging of the external or internal reactor preheaters occurred during either run. In addition, no arsine was detected in the plant area, the hydrotreated product, or during unloading of the catalyst bed.

Second-Stage Hydrotreating

The second-stage hydrotreating of shale oil is required to reduce the high levels of impurities such as nitrogen and oxygen that are still present following first-stage hydrotreating. Sulfur, the other major impurity in shale oil, is rather easily removed at high pressure.

This second-stage hydrotreating pilot plant program included the following two major components:

1. A detailed catalyst evaluation and process variable study.

2. The production of three 100-gallon batches of shale oil hydrotreated to different nitrogen levels.

Process Study

A second-stage hydrotreating program was conducted to accomplish the following objectives:

1. Determination of the optimum process conditions required to achieve a maximum 1000 ppm nitrogen level product.

2. Evaluation of a series of catalysts and selection of the most promising of these catalysts for a second-stage hydrotreating feed preparation run.

Catalysts -- The commercial proprietary UOP catalysts employed in this study include three categories of catalysts. The hydrotreating catalysts were designated DSA and DSB; the hydrocracking catalysts were DCA, DCB, DCC, and DCD; the vacuum bottoms hydrocracking catalyst was BCA. This selection of catalysts represents a wide range of catalytic chemistry and was incorporated in the study to determine the best practical catalyst choice.

Equipment and Procedures -- A high pressure pilot plant with a trickle bed reactor holding up to 200 mL of catalyst was employed. For hydrotreating, the plant arrangement consisted of a single block heater, two high pressure separators and a product stripper (Figure 22).

Charge stock, hydrogen rich recycle gas and make-up hydrogen were fed downflow over the catalyst bed. The reactor effluent was sent to a high pressure separator for gas and liquid separation. The bottoms product was sent to a stripper for removal of H_2S, NH_3 and H_2O and light C_1-C_4 cracked gaseous products. Water was injected into the gas line from the first separator. Water and gas proceed to a three-phase separator where the water washed recycle gas is removed and sent back to the reactor, any light condensed hydrocarbons are sent to the stripper, and the water removed from the bottom of the separator. The gas from the stripper was measured and analyzed and the liquid product from the bottom of the stripper was weighed.

The processing procedure was as follows: the first-stage hydro-treated shale oil was charged to the reactor, utilizing a presulfided test catalyst at the initial operating condition of 1700 psi (P-P_B), 5.0 SCFB (H/H_B) hydrogen gas recycle, 0.33 LHSV (L/L_B), and 270°F (T-T_B) catalyst maximum. Note that the base conditions used in this study are relative to those required to effect a high level of nitrogen removal in the corresponding petroleum fraction. The reactor heater was adjusted to bring catalyst maximum temperature to the initial 0°F (T-T_B) test condition and line out the plant for a specified time before starting a test period. All line-out and test period times in this work were 10 hours long. Product analyses and distributions were obtained on appropriate test periods.

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Product quality was monitored by measuring the total nitrogen content of the liquid product. A temperature survey was performed on each of the catalyst systems tested. During selected test periods, the effect of varying residence time (space velocity) was tested. The Occidental shale oil feed was utilized as feedstock for testing seven different catalyst systems; two catalysts were evaluated with Paraho shale oil.

Charge Stocks -- The two first-stage hydrotreated shale oils, Occidental and Paraho, produced in the previous study were used in this work. Inspections of these feeds are shown in Tables 25 through 28.

Catalyst Evaluation -- The performance of each of the catalysts in the study was evaluated on its ability to reduce the nitrogen content of the feed. Each of the catalysts was put through a temperature survey from -20 to +50°F $(T-T_B)$. In some cases a space velocity response was also obtained.

Seven catalysts were tested at 1700 psi $(P-P_B)$ and 0.33 LHSV (L/L_B) using the Occidental shale oil. The processing results and operating conditions are given in Tables 32 to 48. Figures 23 and 24 show the product nitrogen content versus temperature. The one point obtained using DSB is not conclusive as to that catalyst's overall activity. A table of nitrogen product content at various catalyst bed temperatures tested is summarized as follows:

Temp., °F $(T-T_B)$	Nitrogen in Effluent, ppm				
	-20	-10	0	+10	+20
Catalyst					
DCA	1,360	900	650	500	400
DCB	1,830	1,130	700	440	270
DCC	4,700	3,700	2,900	2,100	1,700
DCD	~3,100	2,060	1,360	900	585
DSA	2,200	1,400	1,100	950	800
BCA	1,200	980	780	620	490

Overall, catalyst DCA showed the best performance. At temperatures above $10°F$ $(T-T_B)$ the DCB shows activity equal to or better than the DCA. The BCA catalyst activity remains slightly poorer than the DCA catalyst. Each of the remaining catalysts appeared less effective for nitrogen removal than DCA.

Of the two catalysts tested with the Paraho stock at 1700 psi $(P-P_B)$ and 0.33 LHSV (L/L_B), the DCA catalyst appeared to reduce the nitrogen level more effectively. The data shown in Tables 49 to 54 and in Figure 25 are summarized as follows:

Temp., °F $(T-T_B)$	N in Effluent, ppm				
	-10	0	+10	+20	+30
Catalyst					
DCA	2,350	1,850	1,450	1,150	900
DSB	2,940	2,440	2,000	1,670	1,380

As shown in Figure 25, reduction of space velocity to 0.16 LHSV (L/L_B), indicated no change in the original activity rating of the two catalysts. The DSB could meet the 1,000 ppm product nitrogen content at $-30°F$ $(T-T_B)$ and 0.16 L/L_B.

The total nitrogen content of the Paraho stock was much higher than in the Occidental shale oil (1.9 vs. 1.1 wt-%). A temperature increase of 35 to 40°F was required with the DCA catalyst to reach the 1,000 ppm target nitrogen level in the effluent. This difference in temperature requirement translates into an increase in the catalyst requirement of ~ 50%.

Hydrogen Consumption -- A comparison of hydrogen consumption values as a function of catalyst temperature is shown on Figures 26 and 27. Of all the catalyst systems tested with Occidental feed, the DCA had the highest hydrogen consumption at a given temperature. However, because of the catalyst activity differences, the DCA catalyst required only an intermediate consumption of hydrogen to achieve the 1000 ppm product

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nitrogen level. At the 1,000 ppm nitrogen product content, hydrogen consumption for each of the catalysts tested was as follows:

Catalyst	Hydrogen Consumption, SCFB					
	DCA	DCB	DCC	DCD	BCA	DSA
Occidental Shale Oil	820	780	(900)	820	750	860
Paraho Shale Oil	1515	-	-	-	-	-

Both the DCC and DSA catalysts required higher hydrogen consumption while the BCA and DCB required lower hydrogen consumptions. The DCD and DCA catalysts had equal hydrogen consumption requirements.

When processing the Paraho feed to produce a 1,000 ppm nitrogen product, the hydrogen requirement almost doubled to 1,515 SCFB.

Product Yields -- A summary of the product properties is shown in Tables 55 to 68. The changes in product API gravity are shown versus operating temperature in Figures 28 and 29. With the DCA catalyst, a 30°F increase in temperature produced increases of 2.5 numbers in API gravity with Occidental shale oil, and produced a 2.0 increase with Paraho.

With the DCA catalyst, overall cracking of the Occidental shale oil at 0.33 LHSV (L/L_B) to C_1-C_4 and C_5-C_6 light products is less than that experienced with the Paraho shale oil. As indicated by the ranges of product yields given in Figures 30 and 31, at the conditions used to perform this nitrogen removal work for Occidental shale oil, no significant differences were observed in the hydrocracking activity. The difference in the amount of cracking observed between the two shale oils can be attributed to the higher temperature requirement to process the Paraho feed.

Typical yield distributions using the DCA catalyst over a -20 to +30°F (T-T_B) temperature range at 0.33 LHSV (L/L_B) with both feedstocks were:

		Fresh Feed, Wt-%		
		C_1-C_4	C_5	C_6+
Occidental Shale Oil	-20	0.7	0.5	98.4
	+30	2.1	1.0	96.8
Paraho Shale Oil	-20	0.8	0.5	98.4
	+30	3.0	1.4	95.2

Kinetic Studies -- A kinetic analysis was made for the DCA catalyst for nitrogen removal when processing the Occidental and Paraho feeds. As shown in Figure 32, a good first order relationship was found. The activation energies obtained from the analyses shown on Figures 33 and 34 were 21,000 cal/mol for Occidental shale oil and 19,000 cal/mol for Paraho shale oil. These activation numbers are essentially the same.

Catalyst Selection for High Pressure Hydrotreating -- Based on the catalyst temperature requirement, the C_7+ bottoms product yields, and hydrogen consumptions, the DCA catalyst gave the best performance and was selected for large scale production of hydrotreated Occidental and Paraho shale oil product to contain approximately 1000 ppm nitrogen. This product is to be the feedstock for the Phase II shale oil hydrocracking studies.

Conditions for High Pressure Hydrotreating -- Conditions for high pressure hydrotreating of both the Occidental and Paraho shale oils with use of the DCA catalyst are as follows:

	Occidental	Paraho
Pressure, psi $(P-P_B)$	1,700	1,700
LHSV (L/L_B)	0.33	0.33
H_2 Gas Recycle, SCFB (H/H_B)	5.0	5.0
Temp., °F $(T-T_B)$	-15	+25

Pilot Plant Production Run

The objectives of this pilot plant operation were to produce three 100 gallon batches of hydrotreated shale oil. The first 100 gallon batch

was to be hydrotreated Occidental shale oil containing 500 or less ppm nitrogen. The second batch was to be hydrotreated Occidental shale oil containing about 1000 ppm nitrogen. Finally, a third batch of hydrotreated Paraho shale oil containing about 1000 ppm nitrogen was to be produced. The materials were produced for use as feedstock for the hydrocracking process variable studies.

Catalyst -- The DCA catalyst was specifically selected for this production run based on its performance in the catalyst comparisons study. After the plant was pressure tested, the catalyst was prewetted and sulfided with light cycle oil. Upon completion of sulfiding, fresh feed was introduced into the plant.

Charge Stock -- This run used two different charge stocks. The first charge stock was low pressure hydrotreated Occidental shale oil. This charge stock had been hydrotreated in the pilot plant, Production Run No. 1. A complete inspection of this charge stock is presented in Tables 25 and 26. The second charge stock was low pressure hydrotreated Paraho shale oil. This charge stock had also been hydrotreated in the pilot plant, Production Run 2. A complete charge stock inspection is presented in Tables 27 and 28.

Equipment -- The schematic flow diagram of the plant is presented in Figure 35. Feed was introduced with recycle hydrogen into the top of the external reactor preheater. Gas and oil were heated and then entered the top of the reactor. Water was injected at the reactor outlet to inhibit salt formation. Recycle gas was separated from the oil and water in a high pressure separator. Recycle gas from the high pressure separator passed through a condenser. At the outlet of the condenser water was injected to scrub the recycle gas of ammonia. Water was separated from the recycle gas in a hot flash separator. Heat was not applied to the hot flash separator. Water from the hot flash separator was sent to the caustic scrubber. Recycle gas was returned to the top of the external reactor preheater.

Oil and water from the high pressure separator were sent to a three-phase separator. The three-phase or low pressure separator was used to separate oil, water and gas. Gas and water were sent as separate streams to the caustic scrubber. Oil was directed to a distillation column.

The distillation column was used to strip off water and light hydrocarbons. Gas and water from the column overhead were sent to the caustic scrubber. Column bottoms product was collected in 15 gallon drums.

Operations -- This run was divided into three parts. Part one consisted of high pressure hydrotreating Occidental shale oil to produce 100 gallons of product containing $\leq$ 500 ppm nitrogen. The second part consisted of collecting 100 gallons of high pressure hydrotreated Occidental shale oil with a nitrogen content of about 1000 ppm. Part three consisted of hydrotreating Paraho shale oil to produce $\sim$ 100 gallons of product containing $\leq$ 1000 ppm nitrogen.

To achieve a very low nitrogen product, the following operating conditions were selected for part one:

High Pressure Separator Pressure, psi ($P-P_B$)	1700
Feed LHSV (L/L_B)	0.16
Water Injection at the Reactor Outlet	10 vol-% of Feed Rate
Water Injection at the Outlet of the Recycle Gas Condenser	20 vol-% of Feed Rate
Hydrogen Recycle Gas (H/H_B)	5.0
Product Nitrogen Content, ppm	500 or less
Reactor Average Bed Temperature, °F ($T-T_B$)	-10

As shown in Figure 36 after 120 gallons of low nitrogen product were collected, the operating conditions were changed to produce the high nitrogen product. The LHSV was doubled and the reactor average bed temperature was raised 10°F.

After 139 gallons of high nitrogen Occidental shale oil were collected, operating conditions were again changed to process the Paraho shale oil. Feed LHSV was decreased back to the original level [0.16 (L/L$_B$)] and the reactor temperatures were 20°F higher than those utilized at the same LHSV on Occidental feed [+10°F (T-T$_B$)].

Nitrogen content of the product for part one varied as shown in Figure 36. Final nitrogen content of the composited product of low nitrogen high pressure hydrotreated Occidental shale oil was 0.01 wt-% (see Table 69).

For part two of this run, the daily nitrogen contents of the product are shown in Figure 36. Final product analysis of the composited product of high nitrogen, high pressure hydrotreated Occidental shale oil showed a nitrogen content of 0.09 wt-% (see Table 69).

In part three, the daily nitrogen content of the product varied as depicted in Figure 36. The nitrogen content of the composited product of high pressure hydrotreated Paraho shale oil was 0.07 wt-% (see Table 69).

Analyses of the product drums for all three parts of this run are presented in Figure 37. Following an initial period of high activity, very little hydrocracking of the feedstock occurred as shown by the low vol-% of 550°F minus material in the product. The Bromine Number of the products was observed to vary with the nitrogen content. Hydrotreating of the Occidental shale oil to a nitrogen content of 0.01 wt-% resulted in the Bromine Number being reduced from 10.5 to 0.5. While hydrotreating of the Occidental shale oil to a nitrogen content of 0.09 wt-% caused the Bromine Number to be reduced from 10.5 to 1.1, hydrotreating of Paraho shale oil to a nitrogen content of 0.07 wt-% caused the Bromine Number to be reduced from 19.4 to 1.0.

The final quantities and nitrogen contents of the products collected during this run are summarized below:

	Quantity, Gallons	Nitrogen Content, wt-%
High Pressure Hydrotreated Occidental Shale Oil	115	0.01
High Pressure Hydrotreated Occidental Shale Oil	133	0.09
High Pressure Hydrotreated Paraho Shale Oil	118	0.07

The complete yield structures for the 0.09 wt-% N (approximately 900 ppm) Occidental shale oil and the 0.07 wt-% N (approximtely 700 ppm) Paraho shale oil products are given in Tables 70 and 71, respectively.

After 26 days-on-stream, the plant was shut down and the catalyst flushed with light cycle oil. The catalyst was then removed from the plant in five equal sections and analyzed. As shown in Table 72, a relatively small amount of arsenic was found on the catalyst -- 0.02 wt-% at the top of the bed and 0.004 wt-% at the bottom. Carbon and sulfur were the main deposit constituents, each at about 6-7 wt-%, and were present fairly uniformly throughout the bed. Iron was found in amounts ranging from 0.5 wt-% at the inlet to 0.2 wt-% at the outlet.

Hydrocracking Section

The UOP shale oil upgrading scheme was devised to maximize the production of jet fuel. The primary conversion tool utilized to meet this objective is the UOP HC Unibon process. This proprietary hydrocracking process is highly flexible and can be operated to maximize other distillates such as diesel or gas oil. The technical and economic details developed for Phase II of the program were based on extensive hydrocracking pilot plant operations.

The objectives of the Phase II hydrocracking pilot plant operation were to establish the advantages of a novel process flow arrangement and to demonstrate required hydrocracking process conditions for the production of aviation jet fuel. In addition, small samples of selected fuels

representative of the products from the proposed flow arrangement were to be produced for testing.

The shale oil hydrocracking process conditions used to demonstrate the advantages of this novel system are compared to base conditions for hydrocracking of petroleum vacuum gas oil. An operating pressure of 150 psi below base was used. Other conditions included a space velocity three times that required for petroleum, a combined feed ratio the same as that required for petroleum, and hydrogen gas recycle 1.2 times the base petroleum operation requirement. These conditions produce an accelerated catalyst deactivation that readily provides relative catalyst or process stability information. Commercial conditions are then determined indirectly, but based on the accelerated test data and the known stability characteristics of the base system.

Equipment and Operation

A schematic flow diagram of the single-stage hydrocracking pilot plant used to establish the base performance is shown in Figure 38. Fresh feed is combined with recycle liquid, recycle gas and make-up hydrogen and charged to the reactor. Gas is recycled from the high pressure separator and separator liquid effluent is charged to a series of fractionators that produce liquid product and a recycle liquid stream. During the hydrocracking operation, reactor temperatures were adjusted to achieve 100% conversion of the hydrotreated shale oil to jet fuel in the operations reported herein. In all of the flow schemes studied, no bottoms product was withdrawn as a net product.

Feedstocks

The properties of the Occidental shale oil are shown in Table 69. The high pressure hydrotreated shale oil with a nominal 1000 ppm nitrogen (the initial target level) prepared as hydrocracker feed is shown in the second column. The measured nitrogen content of a large sample of this material is 900 ppm. The sulfur content is less than 300 ppm, the Bromine Number is 1.1, and the oxygen content has been reduced to 651 ppm.

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A more severe high pressure hydrotreating operation with Occidental shale oil produced a substantially lower nitrogen level product. An analysis of this material is shown in the first column of Table 69. The measured nitrogen content is 100 ppm, oxygen content is 144 ppm and the Bromine Number is less than 0.5.

A similar inspection of Paraho shale oil feedstock is shown on Table 69. A very low nitrogen product was not obtained from Paraho shale oil primarily because of insufficient feedstock to make the run. The high pressure hydrotreated product produced from Paraho shale oil contained 700 ppm nitrogen.

Pilot Plant Hydrocracking Operations

The results of processing the Occidental high nitrogen content, high pressure hydrotreated product are shown in Figures 39 through 44. Three flow schemes were investigated. The reference flow scheme is conventional single-stage hydrocracking. As shown in Figure 39, the start-of-run (SOR) temperature required for 100% conversion to JP-8 is 10°F below that required for petroleum based operations. In this accelerated deactivation test and over a 300-hour operating period, the temperature requirement for 100% conversion increased 62°F.

Figure 40 shows data indicating that a modified flow hydrocracking operation dramatically reduces the catalyst deactivation rate. With an SOR temperature essentially the same as that required for the conventional single-stage operation, the temperature requirement for 100% conversion with the modified flow increased only some 14°F over a 300-hour operating period. Figure 41 shows the results of operating with a third hydrocracking flow scheme, termed parallel-flow hydrocracking. Once again, initial activity is essentially the same as in conventional single-stage operation. The catalyst deactivation rate is essentially the same as seen in modified flow hydrocracking.

Data from all these flow schemes are compared on Figure 42. Based on the catalyst stability advantage of parallel flow hydrocracking and on

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engineering considerations, it was concluded that parallei flow would be the selected flow scheme for the subsequent hydrocracking pilot plant study.

Sulfur Addition

A typical petroleum-derived hydrocracking feed contains about 1.5 wt-% sulfur which is virtually completely converted to hydrogen sulfide at reaction conditions. During the process variable study with the hydrotreated shale oil, it was suspected that the low level of sulfur in the hydrocracker feed (< 300 ppm) could be insufficient to maintain the catalyst in an activated state. As a result, different levels of sulfur (as ditertiary butyl disulfide) were added into the hydrocracker feed. The impact of sulfur addition on catalyst activity was very dramatic and spontaneous. For 10,000 ppm of sulfur addition to the feed, the reactor temperature required for complete conversion was reduced by about 20°F. With this positive result, it was decided that sulfur addition to the feed be used in the pilot plant study and that this effect be investigated in subsequent phases of this contract.

Parallel Flow Hydrocracking

The low nitrogen, high pressure hydrotreated Occidental shale oil was processed in a parallel flow hydrocracking operation. Results are shown in Figure 43. The initial temperature requirement was almost 70°F below that required for hydrocracking the 900 ppm nitrogen feed, and the deactivation rate is only 4°F over a 300-hour operating period.

Based on the available pilot plant results on hydrocracking the high and low nitrogen Occidental hydrotreated shale oil, the two different hydrotreating/hydrocracking severity combinations are quite similar in their overall process requirements. However, it was proven that severe hydrotreating of shale oil to low nitrogen content is feasible in one processing step. In addition, the catalyst stability advantage of hydrocracking low nitrogen feed as compared to processing high nitrogen feed has also been demonstrated.

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After successful demonstration of the parallel flow hydrocracking scheme and based on the pilot plant results, yield estimates were prepared to serve as the basis for design of a commercial parallel flow hydrocracker. The feed to the hydrocracker would be high pressure hydrotreated shale oil containing a maximum of 1000 ppm nitrogen. The basic design conditions are shown in Table 73.

The samples of Occidental and Paraho hydrotreated shale oils containing about 700 ppm nitrogen and spiked with sulfur were processed at commercial conditions. This work was intended to provide: a processibility comparison, confirmation of the design operating conditions, product quality information and 500 mL samples for the USAF laboratory analysis. (The Occidental hydrocracker feed containing $\sim$ 700 ppm nitrogen was obtained from a select portion of the nominal 1000 ppm nitrogen level large scale Occidental high pressure hydrotreating production run.) Feedstock inspections are given in Table 74. Figure 44 shows the reactor temperature requirements for 100% conversion to JP-8. The hydrocracking catalyst stability is apparently very similar for both shale oils after each shale oil has undergone two stages of hydrotreatment. However, Paraho shale oil required an initial temperature approximately 10°F higher than that for the Occidental shale oil.

The analytical inspections of the JP-8 jet fuels produced from both shale oils are given in the first two columns in Table 75. Product properties from hydrocracking Occidental feed conformed to JP-8 jet fuel specifications except for the acid content, but the JP-8 jet fuel from Paraho feed was "off-spec" in freeze point and the copper strip corrosion test. A small decrease in product end point of about 20°F should correct the freeze point difference. The failure of the copper strip corrosion test was due to the presence of free sulfur from oxidation of H_2S in the product during pilot plant sample handling. This handling problem could be corrected by caustic treatment of the product immediately as it is produced to remove H_2S. All the jet fuel samples shown in Table 75, except JP-4 fuel from Paraho shale oil, have been treated to remove free sulfur so that they could meet the copper strip corrosion spec. Therefore, there

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should not be any difficulty in reducing the corrosiveness of Paraho products to specification conditions when the same free sulfur removal treatment is applied.

During the commercial demonstration on Occidental hydrotreated feed, the product end point was also decreased to 520°F to produce JP-4 jet fuel. The inspection of JP-4 jet fuel is presented in the third column of Table 75. It conformed to the U.S. military specification (see Table 75). One additional JP-4 jet fuel sample (3794-90) was also produced during the commercial condition demonstration run. It was obtained from products produced during the JP-8 mode of operation. This JP-4 has a high end point (534 vs. 518°F max.) and a high freeze point (-54 vs. -72°F max.).

Since the high pressure hydrotreated Occidental shale oil contains about 30 vol-% of 550°F minus material (jet fuel boiling range), it was decided to initially fractionate out this portion and then hydrocrack the remaining 550°F plus portion. The JP-8 jet fuels from both the 550°F minus fraction and the hydrocracked 550°F plus fraction would then be combined to produce the final JP-8 product. If successful, the size of the hydrocracker would be reduced. Table 76 shows inspections of both 550°F plus and 550°F minus fractions from high pressure hydrotreated Occidental shale oil. The nitrogen content of the 550°F+ fraction was about 1300 ppm compared to about 900 ppm before fractionation (Table 69). The final blended JP-8 jet fuel analyses are shown in the fifth column of Table 75. The nitrogen content has increased to about 65 ppm as compared with 1.1 ppm in the all-hydrocracked case and its acidity was higher than the specification of 0.015 mg KOH/g. Other properties of this material met the JP-8 jet fuel specifications. JP-4 jet fuel was also produced from hydrocracking the 550°F plus fraction. Its product analyses are given in the sixth column of Table 75 and the only "off-spec" property was again the acid content. This experiment, therefore, verified the feasibility of this process scheme.

Two additional JP-8 jet fuel analyses (Sample Nos. 3794-54C and 3794-55C) from hydrocracking ∿ 900 ppm nitrogen Occidental high pressure

hydrotreated shale oil are also presented in Table 75. The high nitrogen feed at higher space velocity (3 vs. 1.4 L/L$_R$ used for the 700 ppm nitrogen feed) produced lower smoke point product, but it met the alternative JP-8 requirement for both hydrogen and naphthalene contents. Both JP-8 analyses also showed high freeze points (-36°F and -45°F vs. -58°F). The seven jet fuel product analyses based on Occidental feed presented in Table 75 represent the seven 500 mL samples sent to the Air Force for their analysis.

The light and heavy naphtha properties from hydrocracking Occidental and Paraho feeds during JP-8 mode of operation are shown in Table 77. The heavy naphthas all contain less than 0.5 ppm nitrogen, but the sulfur contents of 17-100 ppm are high. These high sulfur values are inconsistent with the nitrogen levels and are no doubt due to recombination of H$_2$S with unsaturated hydrocarbons and the result of free sulfur formation following low levels of oxygen contamination with inadequately stripped product.

The ∿900 ppm nitrogen feed was also processed at commercial conditions to provide a processibility comparison with ∿700 ppm nitrogen feed. The pilot plant results indicated that there was about a 15°F catalyst temperature requirement difference in favor of the lower nitrogen feed. In addition, when the hydrocracking temperature requirement for processing the 550°F plus fraction of Occidental feed was compared with the requirement for processing the total 900 ppm nitrogen feed, the heavy fraction required a 15°F higher catalyst temperature to maintain 100% conversion to JP-8 jet fuel.

Yield distributions from various Phase II hydrocracking work are reported in Table 78. The first column represents the product distribution derived from pilot plant data for hydrocracking ∿900 ppm nitrogen Occidental feed. The yield distributions for hydrocracking ∿700 ppm nitrogen Occidental feed for JP-8 and JP-4 jet fuels are given in the next two columns, respectively. The fourth column shows the yield distributions from hydrocracking the 550°F plus fraction of Occidental feed. The last column gives the yield distributions from hydrocracking high pressure hydrotreated Paraho shale oil to JP-8 product.

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It can be observed in Table 78 that the hydrogen consumption increases progressively from about 900 SCFB for hydrocracking the ∿ 700 ppm nitrogen shale oils to JP-8 jet fuel, then to about 1030 SCFB for hydrocracking to JP-4 fuel, and finally to about 1150 SCFB for hydrocracking the 550°F plus fraction of the shale oil. With one exception, the yield distributions in all the JP-8 operating modes for both Occidental and Paraho shale oils are very similar and show about 80 vol-% of jet fuel produced. The increased operating severity required to convert the 550°F+ feed created the exception. This resulted in a greater conversion to naphtha and lighter products.

The major portion of the hydrocracking pilot plant work was completed under Phase II of the contract. The remaining work will be carried out in Phase III to produce five gallons each of various end point kerosines and diesel fuels for Air Force testing. In addition, since the effect of feed sulfur addition on catalyst activity and stability were not fully quantified in Phase II of the program, it will again be addressed in Phase III.

Based on the hydrocracking pilot plant operations to date, it is concluded that: 1) the advantage of modified flow and parallel flow hydrocracking were demonstrated; 2) for technical reasons, parallel flow hydrocracking is selected over modified flow; 3) operating conditions to achieve maximum jet fuel production have been established; 4) yields, hydrogen consumption and product qualities have been determined for both Occidental and Paraho feeds; 5) JP-4 and JP-8 jet fuel specification can be met; 6) the sensitivity of hydrocracking catalyst stability to feed nitrogen content was determined.

TABLE 1. BATCH VACUUM DISTILLATION AND INSPECTION ANALYSIS OF OCCIDENTAL SHALE OIL RECEIVED FROM THE DEPARTMENT OF ENERGY, RIFLE OIL SHALE FACILITY

Rifle, Colorado
UOP SAMPLE NO. 49-1098

Fraction	Shale Oil	No. 1	No. 2	No. 3	No. 4	No. 5	No. 6	No. 7	No. 8	No. 9	Bottoms	H₂O
Cut Temperature, °F		515	569	617	655	705	756	815	875	974		
Yield, Volume % of Shale Oil as Received		9.3	10.0	10.0	10.0	10.0	10.0	10.1	10.1	10.0	9.8	0.7
Yield, Weight % of Shale Oil as Received		8.5	9.5	9.6	9.8	10.0	10.1	10.3	10.3	10.3	10.8	0.8
Position in Shale Oil, L.V.%		0.7-10.0	10.0-20.0	20.0-30.0	30.0-40.0	40.0-50.0	50.0-60.0	60.0-70.1	70.1-80.2	80.2-90.2	90.2-100.0	0.0-0.7
Analyses of Fractions												
Gravity, °API at 60°F.	23.3	37.5	31.3	29.3	26.7	23.0	21.4	20.9	19.9	18.4	10.0	
Specific Gravity at 60°F	0.9141	0.8373	0.8692	0.8800	0.8944	0.9159	0.9234	0.9285	0.9346	0.9440	1.0000	
Total Sulfur, Wt. %		0.61	0.59	0.55	0.60	0.62	0.60	0.50	0.49	0.55	0.80	
Acid Number, mgKOH/g		0.11	0.48	0.84	0.87	0.98	0.94	0.72	0.88	2.11	1.59	
Bromine Number		37	31	28	26	25	23	21	18	17	18	
F.I.A. Analysis, L.V.%												
Paraffins + Naphthenes		38.8	38.3	35.4	32.2	25.6	23.6					
Aromatics + Olefins		61.2	61.7	64.6	67.8	74.4	76.4					
ASTM Elution, Wt. %												
Nonaromatics								32.9	34.8	28.9	Too	
Aromatics								67.1	65.2	70.2	Heavy	
Arsenic Content, ppm		8	10	13	18	18	20	22	26	45	56	
Chlorine Content, ppm		668	20	15	6	4	9	5	2	2	<1	
Nitrogen Content, Total Wt. %		0.97	1.15	1.23	1.42	1.46	1.53	1.42	1.40	1.57	1.77	
Carbon, Wt %		84.25	84.78	85.47	85.04	85.54	84.77	85.08	86.29	85.88	84.67	
Hydrogen, Wt %		12.64	12.27	12.27	12.09	11.80	11.63	11.72	12.07	11.96	11.25	
Color												
ASTM - D 1500		1.5	L2.0	L3.0	4.5	7.0	8.0	L5.5D11	6.5D11	D8.0D11	D8.0D11	
Aniline Point °F.		100	107	116	120	115	127	141	151	155	Too Dark	
Toluene Insolubles (Total Solids) Wt. %			<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.01	0.01	0.07	
Heptane Insolubles, Wt. %		0.01	0.01	0.01	0.01	0.01	0.02	0.03	0.04	0.07	3.04	
Pentane Insolubles, Wt. %		0.01	0.01	0.01	0.01	0.01	0.02	0.05	0.21	0.41	13.9	
Pour Point °F.		-50	-20	5	30	55	75	95	105	110	175	
B. S. & W., Vol. %			<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	0.2	
ASTM, H₂O, Vol. %	0.7											
Conradson Carbon Residue, Wt. %		<0.01	<0.01	<0.01	<0.01	0.01	0.05	0.08	0.36	1.06	11.60	
Existent Gum, g/100ml. *		0.33	31	48	55	67	87	87	85	92	Too Heavy	
Viscosities												
Kinematic at 100°F., Cs.		1.670	3.060	4.961	8.709	18.68	39.78					
Kinematic at 122°F., Cs.								43.03	91.11	300.5		
Kinematic at 210°F., Cs.		0.777	1.138	1.520	2.090	3.166	4.794	7.384	11.69	24.17		
Universal at 100°F., Secs.		31.3	36.1	42.3	54.4	92.3	185.7					
Universal at 122°F., Secs.								201	423	1395		
Universal at 210°F., Secs.		24.2	27.7	30.3	33.1	36.8	42.0	50.4	65.3	116.5		
Furol at 250°F., Secs.											2420	
Furol at 275°F., Secs.											1080	
Metals by AAS, ppm												
Fe		<0.2	0.6	0.3	<0.2	<0.2	<0.2	<0.2	<0.2	0.3	636	
Ni		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	0.4	100	
V		<0.2	<0.2	<0.2	<0.2	<0.2	<0.2	<0.2	<0.2	<0.2	7.6	
Distillation Type		D86	D86	D86	UI	D1160	D1160	D1160	D1160	D1160	D1160	
I.B.P. °F		330	442	500	554	550	605	660	725	730	890	
5%		358	452	512	564	632	685	743	810	855	1005	
10		370	459	518	569	642	697	755	816	875	1028	
20		386	467	525	576	650	704	763	821	887	1030	
30		397	474	531	581	657	711	770	827	895		
40		407	481	536	585	663	718	777	833	902		
50		418	488	542	590	670	726	783	838	912		
60		429	496	550	596	678	735	790	844	923		
70		440	505	558	603	687	745	798	851	935		
80		453	518	567	612	697	758	808	860	948		
90		473	540	584	630	715	775	826	878	965		
95		489	555	600	645	737	791	841	898	986		
E.P. °F.		518	585	632	675	771	822	870	929	1030		
% Recovered		99.0	99.0	98.5	98.5	99.0	99.0	99.0	99.0	99.0	20.0	
% Bottoms		1.0	1.0	1.5	1.5	1.0	1.0	1.0	1.0	1.0	80.0	

*Samples outside scope of method. Results probable low due to mechanical loss. Use with caution.

TABLE 2. BATCH VACUUM DISTILLATION AND INSPECTION ANALYSES OF PARAHO SHALE OIL RECEIVED FROM THE DEPARTMENT OF ENERGY, ANVIL POINTS FACILITY, RIFLE, COLORADO

UOP Sample No. 39-706

Fraction	Shale Oil	No. 1	No. 2	No. 3	No. 4	No. 5	No. 6	No. 7	No. 8	No. 9	Bottoms	H₂O
Cut Temperature, °F		495	580	651	710	775	831	890	945	1026		
Yield Volume % of Shale Oil as Received		9.7	9.9	9.9	10.0	10.0	10.0	10.0	10.0	10.0	10.2	0.3
Yield, Weight % of Shale Oil as Received		8.6	9.3	9.6	9.9	10.0	10.1	10.3	10.4	10.5	11.0	0.3
Position in Shale Oil, L.V.%		0.3-10.0	10.0-19.9	19.9-29.8	29.8-39.8	39.8-49.8	49.8-59.8	59.8-69.8	69.8-79.8	79.8-89.8	89.8-100.0	0.0-0.3
Analyses of Fractions												
Gravity, °API at 60°F	20.6	39.3	30.5	26.0	21.9	19.9	18.8	16.9	14.9	13.7	9.3	
Specific Gravity at 60°F	0.9303	0.8285	0.8735	0.8984	0.9224	0.9346	0.9415	0.9535	0.9665	0.9745	1.0050	
Total Sulfur, wt. %		0.80	0.79	0.74	0.78	0.69	0.64	0.69	0.66	0.70	0.57	
Acid Number, mgKOH/g		1.66	4.60	4.10	2.36	3.59	0.90	0.61	0.50	0.26	0.009	
Bromine Number		53	44	42	41	39	35	30	25	27	26	
F.I.A. Analysis, L.V.%												
Paraffins + Naphthenes		30.8	21.2	18.9	13.0							
Aromatics + Olefins		69.2	78.8	81.1	87.0							
ASTM Elution, wt.%												
Nonaromatics						20.4	22.0	21.5	14.5	26.6	Too	
Aromatics						79.6	78.0	78.5	85.5	73.4	Heavy	
Arsenic Content, ppm		6	13	15	21							
Chlorine Content, ppm		477	22	24	6	25	19	3	2	3	<1	
Nitrogen Content, Total, wt.%		0.99	1.38	2.08	2.17	2.31	2.14	1.99	2.32	2.61	3.14	
Carbon, Wt.%		84.00	84.43	84.31	83.95	84.60	84.97	83.77	84.69	85.21	85.06	
Hydrogen, Wt.%		12.50	11.83	12.03	11.30	11.32	11.29	11.13	11.14	11.56	10.29	
Color												
ASTM - D 1500		L2.0	L4.5	L7.5	14.0 Dil	L6.5 Dil	D8.0 Dil	D8.0 Dil	D8.0 Dil	D8.0 Dil	D8.0 Dil	
Aniline Point °F*		84	87	92	85	90	97	101	102	98	Too Dark	
Toluene Insolubles (Total Solids) wt.%			<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.01	0.03	0.09	
Heptane Insolubles, wt.%		<0.01	<0.01	0.01	0.06	0.12	0.12	0.17	0.20	0.25	5.60	
Pentane Insolubles, wt.%		<0.01	<0.01	0.15	0.22	0.23	0.42	0.61	0.86	1.70	13.42	
Pour Point °F		-65	-10	25	55	80	100	105	105	100	100	
B. S. & W., Vol. %		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	0.1	
ASTM, H₂O, Vol. %	0.3											
Conradson Carbon Residue, wt. %		<0.01	0.02	0.05	0.15	0.29	0.42	0.98	2.22	5.27	18.0	
Existent Gum, g/100 ml**		0.048	0.058	25	59	72	78	70	38	60	Too Heavy	
Viscosities												
Kinematic at 100°F, Cs.		1.401	3.272	7.447	18.45	47.18						
Kinematic at 122°F, Cs.							53.71	119.6	319.7	807.7		
Kinematic at 210°F, Cs.		0.694	1.176	1.886	3.124	5.236	8.365	13.31	24.09	43.55		
Universal at 100°F, Secs.		29.6	36.8	50.3	91.4	220						
Universal at 122°F, Secs.							250	555	1484	3750		
Universal at 210°F, Secs.		23.1	28.0	32.4	36.6	43.4	53.6	71.4	115.8	204		
Furol at 140°F, Secs.											7695	
Furol at 210°F, Secs.											290	
Metals by AAS, ppm												
Fe		≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	0.7	0.6	1.8	526	
Ni		≤0.05	≤0.05	≤0.05	≤0.05	≤0.05	≤0.05	≤0.05	≤0.05	0.8	28	
V		≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	≤0.2	2.7	
Distillation Type		D 86	D 86	U 1	U 1	D 1160	D 1160	D 1160	D 1160	D 1160	D 1160	
I.B.P. °F		300	435	514	570	602	672	735	760	764	959	
5%		325	447	526	590	680	757	811	862	950	1030	
10		340	454	535	600	698	765	819	873	985	1040	
20		356	465	547	610	706	771	826	880	1002	1045	
30		372	475	555	618	715	777	831	886	1003	1058	
40		384	484	561	626	723	783	837	892	1005	1075	
50		396	493	568	634	730	790	842	900	1007		
60		410	503	576	642	737	797	848	909	1010		
70		424	513	584	650	745	804	855	920	1017		
80		440	525	594	661	760	816	866	932	1028		
90		461	548	611	679	789	839	888	954	1045		
95		480	568	626	691	810	855	910	975	1067		
E.P. °F		520	608	654	715	850	895	958	1010	1100		
% Recovered		98.5	98.5	98.5	98.5	99.0	99.0	99.0	99.0	99.0	40.0	
% Bottoms		1.5	1.5	1.5	1.5	1.0	1.0	1.0	1.0	1.0	60.0	

*Approximate

**Samples outside scope of method, Results probably low due to mechanical loss. Use with caution

TABLE 3. OCCIDENTAL SHALE OIL FEED PREPARATION RUN NO. 1
(Desalting)

Operating Conditions

Charge Rate, gph	45
Pressure, psig	85
Temperature, °F	225

Stage 1:

Water Injection Rate, cc/min	4.0
Current, milliamps	67
Voltage, volts	9,240
Temperature, °F	225

Stage 2:   Off

Plant Operation Mode:	Once-through
	Single-Stage
Total Quantity Processed:	1,250 gallons

TABLE 4. OCCIDENTAL SHALE OIL FEED PREPARATION RUN NO. 2
(Desalting)

Operating Conditions

Charge Rate, gph	45
Pressure, psig	90
Temperature, °F	224

Stage 1:

Water Injection Rate, cc/min	113
Current, milliamps	59
Temperature, °F	230

Stage 2:

Water Injection Rate, cc/min	113
Current, milliamps	63
Temperature, °F	233

Plant Operation Mode:	Once-through Two-Stage
Total Quantity Processed:	2,450 gallons

TABLE 5. OCCIDENTAL SHALE OIL FEED PREPARATION

Analytical Inspections

	Raw Oil	Desalted Oil	Desalted Oil	Blend of Desalted Oil
Feed Prep. Run No.	-	1	2	1 and 2
Quantity, gallons	-	1,250	2,450	3,700
API Gravity at 60°F	23.3	23.5	22.8	22.9
Specific Gravity at 60°F	0.9141	0.9129	0.9172	0.9165
Distillation (D-1160), °F				
IBP	402	400	402	376
5%	489	489	496	467
10%	521	520	526	570
30%	602	623	620	670
50%	707	728	721	712
70%	810	-	823	820
90%	946	-	-	953
95%	991	-	-	-
EP	-	-	-	-
% Over	95			87
Pour Point, °F	+75	+60	-	+65
Viscosities				
Kinematic at 122°F, cst	20.28	20.56	20.66	21.94
Kinematic at 210°F, cst	5.038	5.181	4.927	5.268
Carbon, wt-%	84.5	84.1	85.2	84.85
Hydrogen, wt-%	12.2	12.5	11.7	12.27
Total Nitrogen, wt-%	1.4	1.4	1.3	1.51
Sulfur, wt-%	0.64	0.63	0.63	0.64
Chloride, wt-ppm	< 1.0	< 1.0	< 1.0	< 1.0
BS and W, vol-%	1.0	0.1	-	0.2
Conradson Carbon, wt-%	1.17	1.15	1.19	1.36
Ash, wt-%	0.008	0.004	0.008	0.014
Heptane Insolubles, wt-%	0.43	0.60	0.23	0.34
Acid No. Mg KOH/g spl.	1.29	-	-	-
Total Solids, wt-%	0.17	0.05	-	-
C_5 Insolubles, wt-%	1.00	1.15	1.81	1.65
Metals by Emission				
Fe, ppm	31	50	74	45
Ni, ppm	5.6	7.3	11	6.7
V, ppm	0.42	0.44	0.93	0.42
Pb, ppm	< 0.1	0.16	< 0.1	< 0.1
Cu, ppm	0.07	0.09	0.30	< 0.1
Na, ppm	11.0	3.2	13	11
Mo, ppm	1.8	2.7	5.1	1.6
Arsenic, wt-ppm	26.0	25.0	27.0	27.5
Bromine No.	29.8	28.6	-	23.6
Diene Value	30.3	27.2	-	-
Oxygen, wt-%	-	-	-	0.65
Water, wt-%	-	-	-	0.05

TABLE 6. PARAHO SHALE OIL FEED PREPARATION

RUN NO. 3

Operating Conditions

Charge Rate, gph	45
Pressure, psig	95
Temperature, °F	222

Stage 1:

Water Injection Rate, cc/min	108
Current, milliamps	89
Temperature, °F	222

Stage 2:

Water Injection Rate, cc/min	108
Current, milliamps	100
Temperature, °F	222

Plant Operation Mode:	Once-through
	Two-Stage
Total Quantity Processed:	400 gallons

TABLE 7. PARAHO SHALE OIL FEED PREPARATION

Analytical Inspections

	Raw Oil	Desalted Oil
API Gravity at 60°F	19.9	19.8
Specific Gravity at 60°F	0.9346	0.9352
Distillation (D-1160), °F		
IBP	407	408
5%	498	471
10%	533	525
30%	683	670
50%	802	789
70%	904	889
90%	1083	1005
95%	-	-
EP	-	-
% Over	90	90
Pour Point, °F	+85	+85
Viscosities		
Kinematic at 122°F, cst	24.27	24.84
Kinematic at 210°F, cst	5.268	5.388
Universal at 122°F, sus	116.4	118.7
Universal at 210°F, sus	43.5	43.9
Carbon, wt-%	84.6	85.5
Hydrogen, wt-%	11.4	11.96
Nitrogen, wt-%	2.1	2.19
Sulfur, wt-%	0.65	0.70
Chloride, wt-ppm	< 1.0	< 1.0
BS and W, vol-%	< 0.1	< 0.1
Conradson Carbon, wt-%	2.60	2.71
Ash, wt-%	0.012	0.007
Heptane Insolubles, wt-%	0.71	0.47
Acid No. Mg KOH/g spl.	1.80	-
Total Solids, wt-%	0.04	0.03
C_5 Insolubles, wt-%	1.39	1.25
Metals by Emission		
Fe, ppm	38	38
Ni, ppm	2	2.2
V, ppm	0.1	0.3
Pb, ppm	< 0.1	< 0.1
Cu, ppm	0.05	0.08
Na, ppm	38	1.1
Mo, ppm	< 0.01	0.1
Arsenic, wt-ppm	17	19
Bromine No.	41.5	34.7
Total Oxygen, ppm	4500	4500
Water, wt-%	-	< 0.01

TABLE 8. LOW PRESSURE HYDROTREATING

RUN 1 OPERATIONS SUMMARY

Feedstock: Desalted Occidental Shale Oil, UOP No.49-1098A
UOP-DSA Catalyst

		Process Conditions							Conrad.							
												Product Analyses				
Hrs. On Stream	L/Lb LHSV	P-Pb Pres., psig	T-Tb ABT, °F	H/Hb H2 Circ., SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Conrad. Carbon, wt-%	Ash, wt-%	H/C, wt-%	S, wt-%	N, wt-%	O, wt-%	Fe, ppm	As, ppm
Feed					22.9	0.9165	23.6	1.36	0.014	12.27/84.66	0.64	1.513	0.28	42	27
40-48	0.5	0	-300	0.67	23.6	0.9123	22.6	1.05	<0.001	12.58/85.62	0.50	1.52	0.36	42	25
64-72	0.5	0	-150	0.67	24.6	0.9065	21.7	0.80	0.003	12.35/86.03	0.56	1.46	0.43	21	12
136-144	0.5	0	-30	0.67	26.5	0.8956	15.8	0.68	<0.001	12.61/85.98	0.26	1.22	0.40	5.6	6
168-176	0.5	0	+20	0.67	27.2	0.8916	11.5	0.52	0.002	12.71/85.52	0.20		0.146	5.8	2
192-200	0.5	0	+15	1.33	26.3	0.8967	12.6	0.53	<0.001	12.46/85.09				5.7	9

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TABLE 9. LOW PRESSURE HYDROTREATING

RUN 1 OPERATIONS SUMMARY

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A

UOP-DSA Catalyst

| | Process Conditions | | | | | Product Analyses | | | | | | | | | | |
Hrs. On Stream	L/L_b LHSV	$P-P_B$ Pres., psig	$T-T_B$ ABT, °F	H/H_B H_2 Circ., SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Conrad. Carbon, Wt-%	Ash, Wt-%	H/C, Wt-%	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
Feed					22.9	0.9165	23.6	1.36	0.014	12.27/84.99	0.64	1.513	0.28	42	27
216-224	0.5	0	+40	1.33	27.1	0.8922	10.7	0.43	<0.001	12.55/85.09	0.10	1.04	0.423	5.3	3
240-248	0.33	0	+40	1.33	30.0	0.8762	9.0	0.38	0.002	12.64/85.72	0.09	1.06	0.205	2.3	1.2
280-288	0.16	0	+40	1.33	30.2	0.8751	5.8	0.26	<0.001	13.06/86.07	-	0.951	0.251	<1	<1

TABLE 10. LOW PRESSURE HYDROTREATING

RUN 1 OPERATIONS SUMMARY

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A

UOP-DSA Catalyst

	Process Conditions				Product Analyses										
Hrs. On Stream	L/L$_b$ LHSV	P-P$_B$ Pres., psig	T-T$_B$ ABT, °F	H/H$_B$ H$_2$ Circ., SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Conrad. Carbon, Wt-%	Ash, Wt-%	H/C, Wt-%	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
Feed					22.9	0.9165	23.6	1.36	0.014	12.27/84.99	0.64	1.513	0.28	42	27
328-336	0.5	0	-130	1.33	23.7	0.9117	19.6	0.82	0.002	12.46/86.05	0.36	0.80		8.2	8
400-408	0.5	0	-140	1.33	23.8	0.9111	20.6	0.76	<0.001	12.35/85.62	0.42	0.79	0.46	15.1	7
480-488	0.5	0	-130	1.33	23.5	0.9129	22.5	0.94	<0.001	12.30/85.67	0.56	0.753	0.44	21.1	17
640-720	0.5	0	-150	0.67	24.6	0.9065	21.7	0.80	0.003	12.35/86.03	0.56	1.46	0.43	21	12

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TABLE 11. LOW PRESSURE HYDROTREATING

RUN 1 OPERATIONS SUMMARY

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A

UOP-DSA Catalyst

	Process Conditions							Conrad.		Product Analyses					
Hrs. On Stream	L/L_b LHSV	$P-P_B$ Pres., psig	$T-T_B$ ABT, °F	H/H_B H_2 Circ., SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Carbon, Wt-%	Ash, Wt-%	H/C, Wt-%	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
Feed					22.9	0.9165	23.6	1.36	0.014	12.27/84.99	0.64	1.513	0.28	42	27
216-224	0.5	0	+40	1.33	27.1	0.8922	10.7	0.43	<0.001	12.55/85.09	0.10	1.04	0.423	5.3	3
520-528	0.5	200	+40	1.33	27.5	0.8899	10.2	0.33	<0.001	12.7/85.71	0.044	1.13	0.191	2.9	4
568-576	0.5	480	+40	1.33	27.9	0.8877	8.5	0.31	<0.001	12.65/85.05	0.04			3.6	5
624-632	0.5	460	+60	1.33	29.0	0.8816	6.3	0.22	<0.001	12.73/85.68		0.91		2.1	5
680-688	0.5	460	+85	1.33	29.8	0.8772	6.1	0.18	<0.001	12.77/84.76		0.68	0.10	3.1	3
744-752	0.5	460	+100	1.33	30.4	0.8740	5.3			13.00/85.53	0.070	0.721	0.078	<0.1	2
760-768	0.5	460	+95	1.33			4.9								

TABLE 12. LOW PRESSURE HYDROTREATING

RUN 1 OPERATIONS SUMMARY

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A

UOP-DSA Catalyst

| | Process Conditions | | | | Product Analyses | | | | | | | | | | |
Hrs. On Stream	L/L_b LHSV	$P-P_B$ Pres., psig	$T-T_B$ ABT, °F	H/H_B H_2 Circ., SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Conrad. Carbon, Wt-%	Ash, Wt-%	H/C, Wt-%	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
Feed					22.9	0.9165	23.6	1.36	0.014	12.27/84.99	0.64	1.513	0.28	42	27
784-792	0.5	0	-30	0.67			14.4								
808-816	0.5	0	-30	0.67	25.2	0.9030	15.7			12.40/85.46	0.23	1.17		2.2	6
840-848	0.5	0	-30	0.67	25.0	0.9042	15.9			12.13/84.77	0.29				2
136-144	0.5	0	-30	0.67	26.5	0.8956	15.8	0.68	<0.001	12.61/85.98	0.26	1.33		5.6	6

TABLE 13. LOW PRESSURE HYDROTREATING

RUN 2

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A
                          UOP-DSA Catalyst

Conditions: 0 $P-P_B$, 1.33 H/H_B, 0.33 L/L_B

Hours On Stream	Average Bed Temp. $T-T_B$, °F	API 60°F	Sp. Gr. 60/60	Fe, ppm	Arsenic ppm
Feed		22.9	0.9165	42	27
82-90	+30	27.2	0.8894	1.5	1.4
122-130	+30	27.6	0.8894	2.1	3
170-178	+30	27.4	0.8905	4.7	7
218-226	+50				4
226-234	+50	28.1	0.8866	1.3	7
234-242	+60				4
290-298	+50	28.1	0.8866	1.9	1
322-330	+50	28.1	0.8866	2.7	4
346-354	+50	28.3	0.8855	2.0	7
378-386	+50	28.3	0.8855	2.5	1
410-418	+50	28.2	0.8860	2.7	1
450-458	+50	28.2	0.8860	2.3	2

TABLE 14. LOW PRESSURE HYDROTREATING
RUN 3 SUMMARY

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A α-Alumina.

Conditions: 0 P-P_B, 0.33 LHSV, 1.33 SCFB Hydrogen.

Hours On Stream	Ave. Bed Temp. T-T_B, °F	API 60°F	Sp. Gr. 60/60	Fe, ppm	Arsenic, ppm	N, Wt-%	S, Wt-%
Feed		22.9	0.9165	42	27	1.51	0.64
43-59	+170	22.6	0.9182	32	29	1.50	0.69
91-107	+70	22.8	0.9170	25	24	1.48	0.65
139-155	+25	23.0	0.9159	20	16	1.11	0.56
163-171	+50	23.3	0.9141	17	18	-	-
187-227	+50	23.3	0.9141	17	16	-	0.33
243-251	+50	23.2	0.9147	19	16	1.51	0.52

TABLE 15. PRODUCT ANALYSES

RUN 3 SUMMARY (Continued)

Feedstock: Desalted Occidental Shale Oil, UOP No. 49-1098A α-Alumina.

Conditions: 0 P-P$_B$, 0.33 LHSV, 1.33 SCFB Hydrogen.

Hours On Stream	C$_7$ Insol., Wt-%	Br. No.	C	H
243-213	0.10	20.8/21.6	85.42	12.12

TABLE 16. LOW PRESSURE HYDROTREATING

RUN 4 OPERATIONS SUMMARY

Feedstock: Desalted Paraho Shale Oil, UOP No. 39-706A

UOP-DSA Catalyst

	Process Conditions				Product Analyses												
Hrs. On Stream	L/L$_B$ LHSV	P-P$_B$ Pres., psi	T-T$_B$ ABT, °F	H/H$_B$ H$_2$ Circ. SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Con. Carbon Wt-%	C$_7$ Insol., Wt-%	H/C, Wt-%	Visc. at 122°F	Pour Pt., °F	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
Feed:					19.8	0.9352	34.7	2.71	0.47	11.86/88.29			0.70	2.19	1.58	38	19
10-50	0.5	0	-300	0.67	20.0	0.9346				11.41/84.10			0.70	2.26	0.41	24	13
50-98	1.50	0	-20	0.67	22.9	0.9165	23.8	1.33	0.11	11.89/84.04	19.62 cst 96.33 sus	85	0.38	2.10	0.20	2.8	2
98-130	1.50	0	+20	0.67	25.5	0.9013	19.5			12.10/84.67	16.32 cst 82.72 sus	90	0.27	1.97	0.088	<1	2
130-162	1.52	0	+20	1.33	24.9	0.9042	19.6			12.06/84.90	17.09 cst 85.92 sus	90	0.14	1.81	0.173	<1	1
162-194	1.49	0	+50	1.33	24.3	0.9082	16.6	1.55	0.02	12.21/85.01	14.99 cst 77.51 sus	85	0.12	1.79	0.109	<1	<1
194-218	1.03	0	+50	1.33	26.9	0.8933	14.9	0.97	0.08	12.22/84.83	13.11 cst 70.30 sus	85	0.07	1.72	0.260	<1	<1
218-266	0.52	0	+40	1.33	26.7	0.8944	11.8			12.37/85.80			0.06	1.48	0.066	<1	<1
266-298	1.49	+200	+50	1.33	25.2	0.9030	15.6			10.32/85.51			0.10	1.70	0.177	2.1	<1

TABLE 17. LOW PRESSURE HYDROTREATING

RUN 4 OPERATIONS SUMMARY (Continued)

Feedstock: Desalted Paraho Shale Oil, UOP No. 39-706A

UOP-DSA Catalyst

| | Process Conditions | | | | | Product Analyses | | | | | | | | | | | | |
| | L/L_B | $P-P_B$ | $T-T_B$ | H/H_B | | | | | Con. | C_7 | | | Pour | | | | | |
Hrs. On Stream	LHSV	Pres., psi	ABT, °F	H_2 Circ. SCFB	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Carbon Wt-%	Insol., Wt-%	H/C, Wt-%	Visc. at 122°F	Pt., °F	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
298-330	1.48	+450	+50	1.33	26.4	0.8961	13.9			12.30/85.00			0.07	1.63	0.771	2.0	<1
330-362	1.51	+450	+90	1.33	29.1	0.8811	11.1			12.57/84.89			0.14	1.37	0.308	1.2	<1
362-394	1.49	+450	+110	1.33	29.9	0.8767	9.9			12.61/85.72			0.05	1.18	0.067	1.4	<1
394-442	1.49	+450	-20	1.33	23.8	0.9111	22.3			12.09/84.27			0.23	1.97	0.194	2.0	<1
442-506	1.52	0	-20	1.33	24.3	0.9082										2.7	1

TABLE 18. LOW PRESSURE HYDROTREATING

RUN 5 OPERATIONS SUMMARY

Feedstock: Deashed Occidental Shale Oil, UOP No. 39-706A

UOP-DRA Catalyst

	Process Conditions									Product Analyses							
Hrs. On Stream	L/L$_B$ LHSV	P-P$_B$ Pres., psi	T-T$_B$ ABT, °F	H/H$_B$ H$_2$ Circ., SCF/B	API 60/60°F	Sp. Gr. 60/60°F	Bromine No.	Con. Carbon Wt-%	C$_7$ Insol. Wt-%	H/C, Wt-%	Visc. at 122°F	Pour Pt., °F	S, Wt-%	N, Wt-%	O, Wt-%	Fe, ppm	As, ppm
Feed:					22.9	0.9165	23.6	1.36	0.014	12.27/84.99	26.0 cst 173.8 sus	45	0.64	1.51	0.28	42	27.5
12-60	0.5	0	-30	0.67	24.7	0.9059	13.7			12.23/85.03	17.61 cst 87.9 sus	40 40	0.26	1.39	0.424	13	5
60-101	0.5	0	-45	1.33	24.4	0.9076	15.0			12.24/85.80	18.85 cst 93.1 sus	45 sus	0.22	1.41	0.456	8.7	5
101-141	0.5	0	+40	1.33	26.3	0.8967	10.8	0.45	0.03	12.56/85.93	12.86 cst 69.3 sus	45	0.09	1.18	0.251	1.8	1
141-182	0.33	0	+40	1.33	27.0	0.8927	8.8	0.41	0.02	12.56/86.25	12.09 cst 66.4 sus	55	0.05	1.11	0.315	1.5	1
182-222	0.5	0	-40	1.33	24.3	0.9082	14.1			12.33/85.50	18.75 cst 92.6 sus	55	0.24	1.61	0.364	12.3	3

TABLE 19. BLENDED PRODUCT
FIRST STAGE HYDROTREATING

Occidental Shale Oil, DSA Catalyst

RUN 2

API Gravity at 60°F	28.2
Sp. Gr. at 60°F	0.8860

Distillation (D-1160), °F

IBP	228
5%	420
10%	464
20%	520
30%	569
40%	620
50%	680
60%	739
70%	803
80%	872
90%	959
92%	983
% Over	92.0
% Botts.	8.0

Sulfur, wt-%	< 0.05
Nitrogen, wt-%	1.06

Pour Point, °F	+65
Con. Carbon, wt-%	0.36
C_7 Insolubles, wt-%	0.01
Bromine Number	8.3

Viscosity at 122°F, cst	9.450
Viscosity at 210°F, sus	56.98

Arsenic, ppm	1.0
Iron, ppm	2.5
Carbon, wt-%	85.37
Hydrogen, wt-%	12.66

TABLE 20. ANALYSES OF USED CATALYST

RUN 1

Component	Iron	Arsenic	Sulfur	Carbon	Nitrogen
			---Wt-%---		
Preheater, α-Al$_2$O$_3$					
Top	0.95	0.093	0.64	0.34	0.05
Bottom	0.80	0.080	0.58	----	0.05
Catalyst (UOP-DSA)					
Top	2.04	1.83	7.10	6.33	0.46
Top Middle	0.82	0.827	6.96	5.63	0.42
Middle	0.78	0.579	6.93	5.92	0.44
Bottom Middle	0.60	0.519	6.81	6.10	0.44
Bottom	0.52	0.403	7.10	5.35	0.46

TABLE 21. ANALYSES OF USED CATALYST

RUN 2

Component	Iron	Arsenic	Sulfur	Carbon	Nitrogen
			---Wt-%---		
Catalyst (UOP-DSA)					
Top	2.28	1.48	5.99	8.80	0.42
Top Middle	1.12	0.66	6.04	8.29	0.41
Middle	0.49	0.22	6.53	8.79	0.40
Bottom Middle	0.25	0.09	6.65	7.70	0.35
Bottom	0.16	0.04	6.75	6.50	0.35

TABLE 22. ANALYSES OF USED CATALYST

RUN 3

Component	Iron	Arsenic	Sulfur	Carbon	Nitrogen
			---Wt-%---		
Preheater, α-Al$_2$O$_3$					
Top	0.47	0.0648	0.26	0.228	<0.05
Bottom	0.37	0.0476	0.16	0.140	<0.05
Catalyst Bed, α-Al$_2$O$_3$					
Top	0.12	0.0127	0.05	0.063	<0.05
Bottom	0.094	0.0096	0.05	0.050	<0.05

TABLE 23. ANALYSES OF USED CATALYST

USED CATALYST

RUN 4

Component	Iron	Arsenic	Sulfur	Carbon	Nitrogen
			Wt-%		
Preheater, α-Al$_2$O$_3$					
Top	0.38	0.052	0.43	-----	0.27
Bottom	0.23	0.067	0.18	-----	0.24
Catalyst (UOP-DSA)					
Top	4.85	1.89	7.66	8.63	0.657
Top Middle	2.36	0.596	7.19	8.30	0.567
Middle	0.79	0.180	7.40	7.48	0.541
Bottom Middle	0.44	0.132	7.25	7.33	0.502
Bottom	0.28	0.068	7.13	7.15	0.476

TABLE 24. ANALYSES OF USED CATALYST

RUN 5

Component	Iron	Arsenic	Sulfur	Carbon	Nitrogen
			----Wt-%----		
Preheater, α-Al$_2$O$_3$					
Top	0.56	0.067	0.41	5.94	0.159
Bottom	0.31	0.057	0.23	4.69	0.105
Catalyst (UOP-DRA)					
Top	1.57	1.28	4.51	10.14	0.765
Top Middle	0.87	0.518	4.04	9.12	0.640
Middle	0.68	0.231	4.40	8.16	0.563
Bottom Middle	0.51	0.128	4.98	7.32	0.626
Bottom	0.48	0.098	5.05	7.35	0.568

TABLE 25. FIRST-STAGE HYDROTREATING

First-Stage Hydrotreated Product

Occidental Shale Oil

API Gravity, at 60°F	26.6	Sulfur (LECO), wt-%	0.03
Specific Gravity at 60°F	0.8951	Oxygen, wt-%	0.20
Distillation (D-1160), °F		Nitrogen, wt-%	1.10
IBP	419	Conradson Carbon, wt-%	0.51
5%	499	Ash, wt-%	0.001
10%	527	Heptane Insoluble, wt-%	0.02
30%	615		
50%	705	Metal by Emission (ppm)	
70%	808	Fe	13
90%	950	Ni	3.6
95%	1015	V	0.29
EP	-	Pb	< 0.1
% Over	95	Cu	0.24
		Na	4.4
Pour Point, °F	+ 55	Mo	0.46

Viscosities

Kinematic at 122°F (cst)	12.94
Kinematic at 210°F (cst)	3.77
Universal at 122°F (sus)	69.6
Universal at 210°F (sus)	38.7
Viscosity Index	

Bromine Number	10.5
Arsenic, ppm	1
Basic Nitrogen, wt-%	0.77

Micro C/H/O/N, Wt-%

Carbon	86.2
Hydrogen	12.7
Oxygen	
Nitrogen	0.2

Table 26. Batch Vacuum Distillation and Inspection Analyses of Hydrotreated Occidental Shale Oil
Received from the UOP Pilot Plants at Riverside, Illinois
for the Department of Energy, Rifle Oil Shale Facility at Rifle, Colorado

UOP Sample No. 49-1098D

Fraction	Hydro-treated Shale Oil	No. 1	No. 2	No. 3	No. 4	No. 5	No. 6	No. 7	No. 8	No. 9	Bottoms
					Overhead Cuts						
Cut Temperature, °F		505	570	605	650	705	760	820	877	965	
Yield Volume % of Hydrotreated Shale Oil as Received		10.0	10.0	10.0	10.0	10.0	10.0	10.0	10.2	10.1	9.7
Yield, Weight % of Hydrotreated Shale Oil as Received		9.3	9.6	9.7	9.8	10.0	10.1	10.2	10.5	10.4	10.4
Position in Hydrotreated Shale Oil, L.V.%		0.0-10.0	10.0-20.0	20.0-30.0	30.0-40.0	40.0-50.0	50.0-60.0	60.0-70.0	70.0-80.2	80.2-90.3	90.3-100.0
Analyses of Fractions											
Gravity, °API at 60°F.	27.1	39.5	33.6	32.0	29.7	26.8	25.1	24.3	23.4	22.7	15.7
Specific Gravity at 60°F.	0.8922	0.8275	0.8571	0.8654	0.8778	0.8939	0.9036	0.9082	0.9135	0.9206	0.9613
Total Sulfur, wt. ppm	64	109	173	210	286	241	217	267	404		
Total Sulfur, wt.%											0.19
											too heavy
ASTM Elution, Wt.%											
Nonaromatics								45.4	48.7	46.5	
Aromatics								54.6	51.3	53.5	
Acid Number, mg KOH/g		0.16	0.07	0.01	0.02	0.02	0.01	<0.01	<0.01	<0.01	0.05
Bromine Number		5.8	9.1	8.9	10.4	10.8	9.8	10.0	8.6	8.1	11.3
F.I.A. Analysis, L.V.%											
Paraffins + Naphthenes		74.3	65.6	62.3	57.2	48.6	42.2				
Aromatics		25.7	34.4	37.7	42.8	51.4	57.8				
Arsenic Content, ppm		<1									
Carbon, wt.%		86.27	86.15	86.01	86.22	86.24	86.00	86.11	86.10	86.39	86.61
Chlorine Content, ppm		9	1.0	1.3	1.6	<1	<1	<1	13	3	7
Nitrogen Content, Total wt.%		0.69	0.92	1.02	1.06	1.18	1.23	1.16	1.73	1.96	1.24
Hydrogen, wt.%		13.51	12.94	12.81	12.77	12.49	12.25	12.59	12.68	12.77	12.28
Color											
ASTM - D 1500		L1.0	L1.5	L1.5	L2.0	L2.5	L3.5	5.0	L7.0	L5.0 D11.	D8.0 D11.
Aniline Point °F.		127.0	132.1	140.9	144.7	142.4	147.0	157.8	174.2	183.9	too heavy
Toluene Insolubles, (Total Solids) wt.%		<0.01	<0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.01	0.05
Heptane Insolubles, wt.%		<0.01	<0.01	<0.01	0.01	0.01	0.01	0.02	0.03	0.03	0.15
Pentane Insolubles, wt.%		0.01	0.01	0.01	0.01	0.02	0.03	0.04	0.06	0.07	1.69
Pour Point °F	-40	-10	10	30	50	70	90	105	115	115	
Molecular Weight		210	239	260	288	325	360	390	484	953	
B. S. & W., Vol. %		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	0.15	
Conradson Carbon Residue, wt.%	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	<0.01	0.01	0.09	6.02	
Existent Gum, g/100ml	0.04	0.10	0.05	0.11	0.06	0.41	21.3	66.5	78.1	100.7	
Viscosities											
Kinematic at 100°F., cSt.	1.709	3.140	4.758	7.609	13.43						
Kinematic at 122°F., cSt.						15.31	27.93	52.44			
Kinematic at 140°F., cSt.									77.95		
Kinematic at 210°F., cSt.	0.800	1.181	1.510	2.002	2.751	3.957	5.868	8.888	16.25		
Universal at 100°F., Secs.	31.4	36.4	41.6	50.8	71.3						
Universal at 122°F., Secs						78.8	132.5	244			
Universal at 140°F., Secs.									362		
Universal at 210°F., Secs.	24.4	28.0	30.1	32.8	35.4	39.3	45.5	54.4	82.9		
Furol at 210°F., Secs.										360	
Furol at 250°F., Secs.										112.3	
Distillation, Type		D86	D86	D86	U1	D1160	D1160	D1160	D1160	D1160	D1160
I.B.P. °F		285	452	505	562	588	621	710	676	778	926
5%		322	465	515	569	618	667	744	797	865	956
10		350	472	520	572	631	688	752	810	875	970
20		388	480	528	578	642	695	761	817	882	1000
30		406	487	534	582	650	701	767	820	888	1030 @ 27%
40		421	494	540	587	655	710	770	824	895	
50		434	502	546	592	661	714	775	828	902	
60		444	510	552	597	668	720	782	833	909	
70		455	518	561	602	679	730	788	838	918	
80		469	529	572	610	690	741	796	848	926	
90		487	546	587	623	710	761	819	860	949	
95		502	560	598	632	724	785	840	880	962	
E.P. °F		532	588	620	650	744	817	869	915	1000	
%Recovered		99.0	98.5	98.5	98.5	99.0	99.0	99.0	99.0	99.0	27.0
%Bottoms		1.0	1.5	1.5	1.5	1.0	1.0	1.0	1.0	1.0	73.0
Emission Spectrographic Analysis of Wet Ash											
Elements	Reported as ppm metal in the oil.										
Fe		0.07	0.13	0.15	0.24	0.19	0.13	0.35	0.28	0.22	130
Ni		0.03	0.03	0.02	0.04	0.03	<0.02	0.02	0.03	0.28	30
V		0.10	0.03	0.04	0.02	<0.02	0.04	<0.02	0.18	0.32	1.9
Pb		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.1	3.2
Cu		0.03	<0.02	<0.02	<0.02	<0.02	0.06	<0.02	<0.02	<0.04	1.9
Na		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.1	34
Mo		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.1	5.0

TABLE 27. FIRST-STAGE HYDROTREATING

First-Stage Hydrotreated Product

Paraho Shale Oil

API Gravity at 60°F	23.0		Sulfur (LECO), wt-%	0.05
Specific Gravity at 60°F	0.9159		Oxygen, wt-%	0.22
Distillation (D-1160), °F			Nitrogen, wt-%	1.943
IBP	395		Basic Nitrogen, wt-%	1.24
5%	477		BS and W, vol-%	0.2
10	521		Conradson Carbon, wt-%	1.08
30	641		Ash, wt-%	< 0.001
50	753		Heptane Insolubles, wt-%	0.32
70	851		Salt as NaCl, lb/M Bbls	1.3
90	993			
% Over	93		Metals by Emission (ppm)	
			Fe	2.1
Pour Point, °F	+ 85		Ni	5.6
			V	0.37
			Pb	0.1
			Cu	0.07

Viscosities

Kinematic at 122°F (cst)	19.62	Na	3.1
Kinematic at 210°F (cst)	4.463	Mo	< 0.1
Universal at 122°F (sus)	96.33		
Universal at 210°F (sus)	40.9	Bromine Number	19.4
Viscosity Index			
		Arsenic, ppm	< 1

Micro C/H/O/N, Wt-%

Carbon	86.2
Hydrogen	12.0
Nitrogen	1.8

FIA, Vol-%

A	Too Hvy
O	Too Hvy
P + N	Too Hvy

Table 28. Batch Vacuum Distillation and Inspection Analyses of Hydrotreated Paraho Shale Oil Received from the UOP Pilot Plants at Riverside, Illinois for the Department of Energy, Anvil Points Facility at Rifle, Colorado

UOP Sample No. 39-7068

Fraction	Hydro-treated Shale Oil	No. 1	No. 2	No. 3	No. 4	No. 5	No. 6	No. 7	No. 8	No. 9	Bottoms
Cut Temperature, °F		505	570	628	685	745	817	887	940	1017	
		10.0	10.0	10.0	10.0	10.1	10.0	10.0	9.9	9.9	10.1

Overhead Cuts

Yield Volume % of Hydrotreated Shale Oil as Received

Yield, Weight % of Hydro-treated Shale Oil as Received		9.1	9.5	9.7	9.9	10.1	10.1	10.2	10.2	10.4	10.8
Position in Hydrotreated Shale Oil, L.V.%		0.0-10.0	10.0-20.0	20.0-30.0	30.0-40.0	40.0-50.1	50.1-60.1	60.1-70.1	70.1-80.0	80.0-89.9	89.9-100.0

Analyses of Fractions

Gravity, °API at 60°F	24.1	38.9	32.4	29.5	26.0	23.3	22.2	21.0	19.1	17.2	13.8
Specific Gravity at 60°F	0.9094	0.8304	0.8633	0.8789	0.8984	0.9141	0.9206	0.9279	0.9396	0.9516	0.9738
Total Sulfur, wt.ppm		183	366	394	296	678	603	792			
Total Sulfur, wt.%									0.06	0.08	0.12
Acid Numbers, mg KOH/g		<0.01	<0.01	<0.01	<0.01	0.25	0.07	0.05	0.61	0.04	0.20
Bromine Number		14.8	17.1	19.4	19.6	21.1	18.2	16.3	17.2	18.5	14.9

F. I. A. Analysis, L.V.%

Paraffins + Naphthenes		70.2	60.1	53.9	42.6	36.5	29.9				
Aromatics		29.8	39.9	46.1	57.4	63.5	70.1				

ASTM Elution, Wt.%

Nonaromatics								35.8	30.0	31.5	too heavy
Aromatics								64.2	70.0	68.5	

Arsenic Content, ppm		<1									
Chlorine Content, ppm		226	<1	<1	1	3	1	<1	14	<1	<1
Nitrogen Content, Total, wt.%		1.16	1.57	1.74	1.89	2.00	1.45	1.37	1.45	1.65	1.93
Carbon, wt.%		85.31	85.58	85.27	85.50	85.69	85.75	85.55	86.40	85.74	86.23
Hydrogen, wt.%		13.12	12.59	12.37	12.13	11.89	11.78	11.86	11.71	11.35	11.02

Color

ASTM - D 1500		L1.0	L1.5	2.5	L4.0	L5.5	7.5	L5.5 Dil.	6.5 Dil.	8.0 Dil.	D8.0 Dil
Aniline Point °F		121.3	122.5	128.3	125.1	122.7	131.4	142.7	138.0	124.9	too Dark.

Toluene Insolubles (Total Solids) wt.%			<0.01	<0.01	<0.01	<0.01	0.01	0.01	0.01	0.02	0.02
Heptane Insolubles, wt.%		<0.01	<0.01	<0.01	<0.01	0.01	0.01	0.01	0.04	0.07	1.30
Pentane Insolubles, wt.%		<0.01	<0.01	<0.01	0.01	0.02	0.04	0.04	0.38	1.46	2.69
Pour Point °F		-40	zero	30	55	75	95	105	115	115	100
Molecular Weight			215	244	274	329	340	399	454	457	511
B. S. & W., Vol. %		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	0.05
Conradson Carbon Residue, wt.%		<0.01	<0.01	0.01	0.02	0.04	0.13	0.28	0.61	1.27	10.17
Existent Gum, g/100 ml		0.30	0.60	0.95	0.94	*	*	*	*	*	

Viscosities

Kinematic at 100°F., cSt.		1.669	3.380	6.224	11.91	26.26					
Kinematic at 122°F., cSt.							30.10	56.04			
Kinematic at 140°F., cSt.									68.04	163.1	
Kinematic at 210°F., cSt.		0.773	1.205	1.737	2.551	3.915	6.023	8.994	14.44	26.14	
Universal at 100°F., Secs.		31.2	37.2	46.3	65.7	124.9					
Universal at 122°F., Secs.							142.2	260			
Universal at 140°F., Secs.									316	757	
Universal at 210°F., Secs.		24.1	28.2	31.6	34.7	39.2	46.6	55.7	75.7	125.2	
Furol at 140°F., Secs.											393
Furol at 210°F., Secs.											40.5

Distillation, Type		D86	D86	D86	U1	D1160	D1160	D1160	D1160	D1160	D1160
I.B.P. °F.		322	443	525	598	501	577	790	841	855	957
5%		348	455	535	603	678	742	801	857	910	993
10		363	463	541	606	690	750	805	857	924	1008
20		383	477	549	611	697	756	810	861	933	1018
30		398	487	555	617	700	761	813	865	939	1025
40		412	496	561	622	705	767	817	870	944	1030
50		424	504	567	627	711	773	820	875	950	1041
60		436	513	573	632	717	780	825	881	957	1061
70		448	522	580	638	723	789	832	890	964	1095
80		460	532	588	646	734	800	842	902	974	
90		480	547	601	660	753	814	858	927	994	
95		495	562	617	674	776	829	875	947	1010	
E.P. °F.		520	592	649	702	810	862	920	974	1040	
%Recovered		98.5	98.5	98.5	98.5	99.0	99.0	99.0	99.0	99.0	70.0
%Bottoms		1.5	1.5	1.5	1.5	1.0	1.0	1.0	1.0	1.0	30.0

Emission Spectrographic Analysis of Wet Ash
Elements Reported as ppm metal in the oil.

Elements		No. 1	No. 2	No. 3	No. 4	No. 5	No. 6	No. 7	No. 8	No. 9	Bottoms
Fe		0.4	0.10	0.13	0.05	0.02	0.07	7.0	0.13	0.13	14
Ni		<0.02	0.02	0.02	<0.02	<0.02	<0.02	0.20	0.11	0.77	48
V		<0.02	0.03	<0.02	0.21	0.04	<0.02	0.29	0.15	0.11	2.7
Pb		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	<0.1	0.22
Cu		<0.02	<0.02	<0.02	<0.02	<0.02	<0.02	0.02	<0.02	<0.4	0.60
Na		<0.05	0.08	0.6	0.15	0.26	<0.05	<0.05	<0.05	2.2	18
Mo		<0.05	<0.05	<0.05	<0.05	<0.05	<0.05	0.05	<0.05	0.1	0.27

*Insufficient Sample

TABLE 29. LOW PRESSURE HYDROTREATER -- PRODUCT YIELDS

Catalyst: DSA

Feedstock: Dewatered Occidental Shale Oil

Yields, Wt-% of Feed

H_2O	0.50
NH_3	0.50
H_2S	0.65
C_1	0.03
C_2	0.06
C_3	0.09
C_4	0.12
C_5	0.15
C_6^+	98.86
Total	100.96

Chemical Hydrogen Consumption = 581 SCFB

TABLE 30. LOW PRESSURE HYDROTREATER -- PRODUCT YIELDS

Catalyst: DSA

Feedstock: Dewatered Paraho Shale Oil

Yields, Wt-% of Feed

H_2O	0.50
NH_3	0.29
H_2S	0.49
C_1	0.03
C_2	0.06
C_3	0.09
C_4	0.12
C_5	0.16
C_6^+	99.52
Total	101.26

Chemical Hydrogen Consumption = 765 SCFB

TABLE 31. FIRST-STAGE HYDROTREATING

Shale Oil, Used Catalyst Summary

	Upper Section	Upper Mid.	Mid. Section	Lower Mid.	Lower Section
Iron, wt-%	6.3	4.5	1.0	0.43	0.40
Arsenic, wt-%	2.39	1.17	0.32	0.120	0.074
Carbon, wt-%	10.40	10.52	13.38	13.31	14.12
Nitrogen, wt-%	0.47	0.339	0.573	0.590	0.668
Sulfur, wt-%	9.28	8.61	8.22	7.81	7.53
Hydrogen, wt-%	1.6	1.39	1.56	1.37	1.34

TABLE 32. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 1

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCA

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_B)$	Process Conditions		
				H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$	LHSV (L/L_B)
60–70	7	804	1700	5.0	-10	0.33
80–90	9	822	1700	5.0	-10	0.33
150–160	16	781	1700	5.0	-25	0.33
170–180	18	772	1700	5.0	-25	0.33
240–250	25	892	1700	5.0	+10	0.33
260–270	27	855	1700	5.0	+10	0.33
330–340	34	737	1700	5.0	+10	0.50
350–360	36	752	1700	5.0	+10	0.50
420–430	43	782	1700	5.0	-10	0.33
440–450	45	838	1700	5.0	-10	0.33

TABLE 33. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 1 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCA

Hours on Stream	Period	H_2 Consumption, SCFB	Process Conditions				Total N, ppm
			Pres., psi $(P-P_B)$	H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$	LHSV (L/L_B)	
20-30	3	833	1700	5.0	-10	0.33	560
40-50	5	841	1700	5.0	-10	0.33	730
70-80	8	856	1700	5.0	-10	0.33	
110-120	12	807	1700	5.0	-25	0.33	1490
130-140	14	762	1700	5.0	-25	0.33	1550
160-170	17	795	1700	5.0	-25	0.33	
200-210	21		1700	5.0	+10	0.33	430
250-260	26	906	1700	5.0	+10	0.33	
290-300	30	765	1700	5.0	+10	0.5	1305
310-320	32	760	1700	5.0	+10	0.5	1275

TABLE 34. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 1 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCA

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	Viscosity 210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	-	875	264	0.036	86.50	13.39	0.5	0.02	0.35	< 0.01	< 0.5	< 0.5	6.966	2.587	+70	29.7
9	-	865	252	0.038	86.10	13.28	0.5	0.02	0.21	0.03	< 0.5	< 0.5	7.441	2.712	+75	31.2
16	7.7	1645		0.094	86.46	13.36	1.1	0.04	< 0.06	< 0.01	0.6	< 0.5	8.032	2.855	+50	26.3
18	7.6	1510		0.097	86.39	13.33	1.0	0.08	< 0.03	< 0.01	0.5	< 0.05	8.013	2.872	+55	25.7
25	0.70	520		0.019	86.36	13.48	0.5	0.01	0.05	< 0.01	0.5	< 0.05	6.207	2.385	+80	28.0
27	0.10	540		0.025	86.15	13.41	0.5	< 0.01	0.02	< 0.01	< 0.5	< 0.5	2.399	2.399	+80	27.4
34	-	1290		0.075	86.63	13.30	0.9	0.04	0.05	< 0.01	< 0.5	< 0.5	6.683	2.515	+80	33.0
36	-	1327		0.074	86.71	13.28	1.2	0.01	0.07	0.01	< 0.5	< 0.5	6.409	2.431	+80	35.1
43	7.5	1130	401	0.061	86.71	13.28	0.9	0.02	0.06	< 0.01	0.5	< 0.5	7.099	2.619	+80	36.7
45	5.2	1050	366	0.056	86.64	13.44	0.8	0.03	0.04	< 0.01	0.5	< 0.5	7.055	2.612	+80	33.5

TABLE 35. SECOND-STAGE HYDROTREATING

Run 2

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCC

Operating Conditions and Product Properties

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_B)$	Process Conditions H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$	LHSV (L/L_B)
20-30	3	-	1700	5.0	0	0.33
40-50	5	-	1700	5.0	0	0.33
60-70	7	675	1700	5.0	0	0.33
80-90	9	-	1700	5.0	0	0.33
150-160	16	497	1700	5.0	-20	0.33
170-180	18	-	1700	5.0	-20	0.33
240-270	25	-	1700	5.0	+10	0.33
70-80	8	667	1700	5.0	0	0.33
160-170	17	533	1700	5.0	-20	0.50
110-120	12	-	1700	5.0	-20	0.50
130-140	14	-	1700	5.0	0	0.33

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TABLE 36. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 2 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCC

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C5 Insol., Wt-%	C7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	113	3380	-	0.009	86.89	13.21	3.6	0.08	0.01	0.01	< 0.5	< 0.5	7.805	2.787	+85	46.0
9	111	3320	958	0.219	86.97	13.12	1.1	0.06	0.01	0.01	< 0.5	< 0.5	7.825	2.757	+85	47.0
16	119	5370	-	0.366	86.96	13.19	4.9	0.17	< 0.01	0.01	0.7	< 0.5	9.212	3.072	+80	
18	114	5150	1181	0.368	86.68	13.10	4.9	0.14	0.01	< 0.01	< 0.5	< 0.05	9.253	3.133	+80	
25	30	1790	842	0.119	86.99	13.37	2.7	0.04	0.01	0.02	< 1.0	< 0.05	6.577	2.509	+85	36.3
8																
17																
12		5150														
14		4420														
3		2780														
5		3480														

TABLE 37. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 3

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DSA

Hours on Stream	Period	H$_2$ Consumption, SCFB	Process Conditions			
			Pres., psi (P-P$_B$)	H$_2$ Circ., SCFB (H/H$_B$)	Ave. Cat. Temp. (T-T$_B$)	LHSV (L/L$_B$)
60-70	7	805	1700	5.0	0	0.33
80-90	9	-	1700	5.0	0	0.33
190-200	20	822	1700	5.0	+10	0.33
210-220	22	810	1700	5.0	+10	0.33
50-60	6	747	1700	5.0	0	0.33
40-50	5	760	1700	5.0	0	0.33
30-40	4	813	1700	5.0	0	0.33
20-30	3	843	1700	5.0	0	0.33
200-210	21	839	1700	5.0	0	0.33
180-190	19	886	1700	5.0	+10	0.33
110-130	12 & 13	-	1700	5.0	-20	0.33
150-160	16	809	1700	5.0	+10	0.33
170-180	18	875	1700	5.0	+10	0.33

TABLE 38. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 3 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DSA

	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	210°F	Pour Pt., °F	Aromatics, Wt-%
Product Analyses
Period
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	-	1310		0.105	86.39	13.36	0.9	0.02	0.04	0.02	<1	0.9	6.952	2.616	+85	34.8
9	2.5	1430		0.116	86.26	13.27	0.9	0.02	<0.01	0.01	<1	0.5	7.337	2.690	+90	29.5
20	0.1	1080		0.079	86.35	13.50	0.7	<0.01	<0.01	<0.01	<1	<0.5	5.997	2.326	+85	26.8
22	0.1	1010		0.079	86.93	13.48	0.9	<0.01	<0.01	0.01	<1	0.5	6.000	2.323	+85	27.6
6																
5		1140														
4																
3		900														
21																
19																
12 & 13	3.9	2720														
16		820														
18		810														

TABLE 39. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 4

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCC

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_B)$	Process Conditions		
				H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$	LHSV (L/L_B)
60-70	7	-	1700	5.0	0	0.33
80-90	9	640	1700	5.0	0	0.33
180-190	19	-	1700	5.0	+10	0.33
200-210	21	770	1700	5.0	+10	0.33
240-250	25	793	1700	5.0	+30	0.33

TABLE 40. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 4 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCC

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	Viscosity cst 210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	52.7	3020	1154	0.223	86.15	12.98	3.7	0.07	0.04	0.03	1.0	< 0.5	8.019	2.842	+85	45.9
9	67.1	3020		0.223	86.58	13.10	5.8	0.07	0.01	0.02	1	< 0.5	8.044	2.852	+85	48.0
19	10.2	1710		0.111	86.46	13.18	2.7	0.02	0.10	0.02	1	< 0.5	6.570	2.479	+80	34.7
21	16.1	1530		0.099	86.87	13.31	2.5	< 0.01	0.09	< 0.01	1	< 0.5	6.600	2.450	+80	33.5
25	4.3	770	386	0.045	86.50	13.24	1.9	< 0.01	0.02	0.02	t	< 0.05	5.519	2.220	+85	31.1

TABLE 41. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 5

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCB

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_R)$	Process Conditions		
				H_2 Circ., SCFB (H/H_R)	Ave. Cat. Temp. $(T-T_R)$	LHSV (L/L_R)
60-70	7	781	1700	5.0	0	0.33
80-90	9	764	1700	5.0	0	0.33
150-160	16	677	1700	5.0	-10	0.33
170-180	18	664	1700	5.0	-10	0.33
260-270	27	870	1700	5.0	+10	0.33
310-320	32	955	1700	5.0	+35	0.33
350-360	36	1251	1700	5.0	+35	0.16
370-380	38	1329	1700	5.0	+35	0.16
410-420	42	1541	1700	5.0	+25	0.16

TABLE 42. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 5 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst UOP DCB

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C$_5$ Insol., Wt-%	C$_7$ Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	Viscosity cst 210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	-	972	223	0.082	86.68	13.24	1.3	< 0.01	0.16	0.03	< 1	< 0.5	6.268	2.700	+80	
9	6.4	939	250.3	0.075	86.41	13.39	1.3	< 0.01	0.19	0.04	< 1	< 0.5	7.450	2.703	+80	36.6
16	26.6	1840	250.3	0.15	86.43	13.16	1.8	0.07	0.07	< 0.01	< 1	< 0.5	8.498	2.958	+60	
18	22.7	2040	213	0.17	86.67	13.36	2.0	0.08	0.03	< 0.01	< 1	< 0.05	8.800	3.078	+55	
27	2.7	440	107.6	0.03	86.08	13.34	0.9	0.01	0.04	< 0.01	< 1	< 0.5	6.366	2.453	+95	29.9
32	2.1	102	174.9	0.01	86.76	13.51	0.7	< 0.01	0.01	< 0.01	1	< 0.5	4.345	1.857	+80	24.6
36		1.1		< 1*	85.83	13.83	< 0.5	-	0.02	< 0.01	< 1	-	2.851	1.394	+65	12.3
38		1.1		< 1*	85.75	13.79	< 0.5	-	< 0.01	< 0.01	< 1	-	2.966	1.426	+70	13.1
42		7.6		7*	85.79	13.75	<0.5	< 0.01	0.02	0.01	-	< 0.5	3.173	1.511	+75	17.1

* Indicates ppm Instead of wt-%.

TABLE 43. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 6

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCF

Hours on Stream	Period	H_2 Consumption, SCFB	Process Conditions			
			Pres., psi $(P-P_R)$	H_2 Circ., SCFB (H/H_R)	Ave. Cat. Temp. $(T-T_R)$	LHSV (L/L_R)
60-70	7	719	1700	5.0	0	0.33
80-90	9	748	1700	5.0	0	0.33
140-150	15	862	1700	5.0	+10	0.33

TABLE 44. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 6 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DCF

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	22.0	710		0.14	86.88	13.15	1.8	0.02	0.02	0.02	< 1	< 0.5	7.178	2.774	+75	
9	14.2	1685		0.13	86.40	13.27	1.7	0.03	0.01	< 0.01	< 1	2.3	7.678	2.729	+75	
15	3.0	735		0.05	86.40	13.32	1.0	< 0.01	0.01	< 0.01	< 1	0.8	6.431	2.499	+85	30.5

TABLE 45. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 7

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP BCA

Hours on Stream	Period	H$_2$ Consumption, SCFB	Pres., psi (P-P$_B$)	Process Conditions			LHSV (L/L$_R$)
				H$_2$ Circ., SCFB (H/H$_B$)	Ave. Cat. Temp. (T-T$_B$)		
60-70	7	777	1700	5.0	0		0.33
80-90	9	776	1700	5.0	0		0.33
130-140	14	880	1700	5.0	+10		0.33

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TABLE 46. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 7 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP BCA

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	3.3	910		0.06	86.86	13.46	0.9	< 0.01	0.01	< 0.01	1.0	0.8	7.388	2.670	+85	30.7
9	1.0	940		0.07	86.61	13.40	0.9	< 0.01	0.01	< 0.01	< 1	0.7	7.270	2.641	+85	31.4
14	0.9	580		0.04	86.66	13.45	0.8	< 0.01	< 0.01	< 0.01	1	< 0.5	6.129	2.375	+85	29.4

TABLE 47. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 8

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DSB

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_8)$	Process Conditions		
				H_2 Circ., SCFB (H/H_8)	Ave. Cat. Temp. $(T-T_8)$	LHSV (L/L_8)
60-70	7	921	1700	5.0	0	0.33

TABLE 48. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 8 (Continued)

Feed: Hydrotreated Occidental Shale Oil

Catalyst: UOP DSB

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	500	11,500	2000	0.77	86.03	12.60	9.1	0.55	0.21	0.02	0.6	13	14.98	4.180	+65	
7	6.5	700	143	0.05	86.83	13.54	0.9	< 0.01	0.03	0.01	1.0	< 0.5	7.185	2.638	+80	28.5

TABLE 49. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 9

Feed: Hydrotreated Paraho Shale Oil

Catalyst: UOP DCA

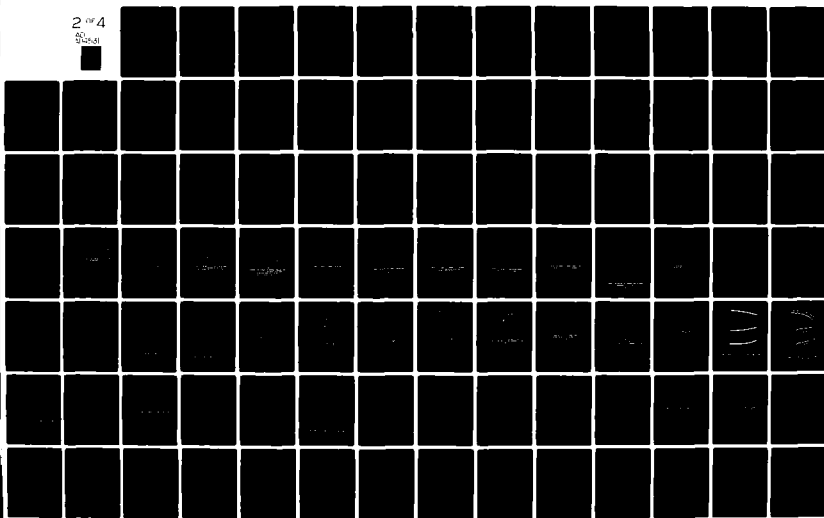| Hours on Stream	Period	H_2 Consumption, SCFB	Process Conditions			
			Pres., psi $(P-P_B)$	H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$	LHSV (L/L_B)
60-70	7	1389	1700	5.0	0	0.33
80-90	9	-	1700	5.0	0	0.33
150-160	16	1428	1700	5.0	+10	0.33
170-180	18	-	1700	5.0	+10	0.33
240-250	25	-	1700	5.0	+10	0.16
260-270	27	2590	1700	5.0	+10	0.16
330-340	34	-	1700	5.0	+35	0.33
350-360	36	-	1700	5.0	+35	0.33
420-430	43	-	1700	5.0	+30	0.16
440-450	45	-	1700	5.0	+30	0.16

TABLE 50. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 9 (Continued)

Feed: Hydrotreated Paraho Shale Oil

Catalyst: UOP DCA

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C_5 Insol., Wt-%	C_7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	946	18,130	2200	1.24	85.00	11.94	17.1	1.04	0.12	0.13	1	< 0.5	17.37	4.463	+90	
7		1880		0.17	86.79	13.31	2.2	0.02	< 0.01	0.01	1	< 0.5	6.640	2.567	+85	34.0
9	9.6	2660	753	0.20	86.87	13.29	2.4	0.03	< 0.01	< 0.01	< 1	< 0.5	6.671	2.528	+85	33.4
16		1510	440	0.10	86.80	13.46	1.8	< 0.01	< 0.01	< 0.01	< 1	< 0.5	5.478	2.259	+85	31.5
18	2.5	1500	308	0.10	85.97	13.19	1.8	0.01	< 0.01	< 0.01	< 1	< 0.05	5.791	2.294	+85	31.6
25	2.1	210.0	152	0.02	86.12	13.48	0.6	< 0.01	< 0.01	< 0.01	< 1	< 0.05	5.080	2.119	+85	22.6
27	1.4	235.0		0.02	86.68	13.66	0.6	< 0.01	< 0.01	< 0.01	< 1	< 0.5	4.984	2.090	+85	23.3
34	0.9	700	266	0.04	86.09	13.34	1.2	< 0.01	0.01	0.01	1	< 0.5	4.289	1.918	+85	28.9
36	1.7	700		0.04	86.76	13.54	1.4	< 0.01	< 0.01	< 0.01	1	< 0.5	4.340	1.915	+85	27.8
43		21	44	< 0.01	86.14	14.28	< 0.5	< 0.01	-	< 0.01	< 1	< 0.5	-	-	+80	18.6
45		21		< 0.01	86.00	13.93	0.5	< 0.01	-	< 0.01	< 1	< 0.5	-	-	+80	19.1

TABLE 51. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 9 (Continued)

Feed: Hydrotreated Paraho Shale Oil

Catalyst: UOP DCA

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_B)$	Process Conditions			LHSV (L/L_B)	Total N ppm
				H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$			
40-50	5	-	1700	5.0	0		0.33	2130
70-80	8	1323	1700	5.0	0		0.33	-
110-120	12	-	1700	5.0	+10		0.33	1510
160-180	17-18	1490	1700	5.0	+10		0.33	-
200-210	21	2188	1700	5.0	+10		0.16	260
220-230	23	2507	1700	5.0	+10		0.16	220
290-300	30	-	1700	5.0	+35		0.33	600
310-320	32	1550	1700	5.0	+35		0.33	600

TABLE 52. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 10

Feed: Hydrotreated Paraho Shale Oil

Catalyst: UOP DSB

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_B)$	Process Conditions		
				H_2 Circ., SCFB (H/H_B)	Ave. Cat. Temp. $(T-T_B)$	LHSV (L/L_B)
60-70	7		1700	5.0	0	0.33
80-90	9		1700	5.0	0	0.33
150-160	16		1700	5.0	+10	0.33
170-180	18		1700	5.0	+15	0.33
240-250	25		1700	5.0	+10	0.16
260-270	27		1700	5.0	+10	0.16
330-340	34		1700	5.0	+25	0.33
350-360	36		1700	5.0	+25	0.33
410-430	42-43		1700	5.0	+25	0.16
430-450	44-45		1700	5.0	+25	0.16
500-520	51-52		1200	5.0	+10	0.16
520-540	53-54		1200	5.0	+10	0.16

TABLE 53. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 10 (Continued)

Feed: Hydrotreated Paraho Shale Oil

Catalyst: UOP DSB

Hours on Stream	Period	H_2 Consumption, SCFB	Pres., psi $(P-P_B)$	Process Conditions		Ave. Cat. Temp. $(T-T_R)$	LHSV (L/L_R)	Total N ppm
				H_2 Circ., SCFB (H/H_B)				
40-50	5		1700	5.0		0	0.33	2700
110-120	12		1700	5.0		+15	0.33	1800
130-140	14		1700	5.0		+15	0.33	2100

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TABLE 54. SECOND-STAGE HYDROTREATING

Operating Conditions and Product Properties

Run 10 (Continued)

Feed: Hydrotreated Paraho Shale Oil

Catalyst: UOP DSB

Product Analyses

Period	Total S, ppm	Total N, ppm	Total O, ppm	Basic N, Wt-%	C, %	H, %	Br. No.	Con. Carbon, Wt-%	C5 Insol., Wt-%	C7 Insol., Wt-%	As, ppm	Fe, ppm	Viscosity cst 122°F	Viscosity cst 210°F	Pour Pt., °F	Aromatics, Wt-%
Feed:	946	18,130	2200	1.24	85.00	11.94	17.1	1.04	0.12	0.13	1	< 0.5	17.37	4.463	+90	
7	51	3200	1263	0.21	86.23	13.13	2.0	0.02	< 0.01	< 0.01	<1	0.34	7.224	2.674	+85	34.4
9	58	2500		0.23	86.42	13.55	2.8	0.03	< 0.01	< 0.01	1	0.21	5.812	2.625	+85	32.8
16		2200		0.17	86.89	13.53	2.1	< 0.01	< 0.01	< 0.01	1	0.17	5.812	2.314	+85	31.1
18	11.7	2000	669	0.15	86.36	12.94	1.5	< 0.01	0.01	0.01	1	0.081	5.487	2.233	+85	29.8
25	2.4	400	463	0.03	86.51	13.75	< 0.5	< 0.01	< 0.01	0.01	<1	1.2	5.027	2.093	-	21.7
27	2.5	400	492	0.03	86.37	13.58	< 0.5	< 0.01	0.01	< 0.01	<1	1.2	4.800	2.043	+80	22.2
34	2.0	1200		0.08	85.96	13.84	1.4	0.01	< 0.01	0.01	<1	0.11	4.145	1.831	+80	25.6
36		1400	351	0.09	86.71	13.50	1.6	< 0.01	0.01	0.01	<1	0.21	4.422	1.924	+85	26.2
42-43	1.6	200	56	0.01	86.22	13.80	0.6	0.01	0.01	0.01	<1	0.37	3.460	1.609	+85	19.4
44-45		200	112	0.01	86.24	13.58	0.6	0.01	0.01	< 0.01	<1	0.79	3.355	1.574	+80	19.7
51-52	2.4	1700	741	0.13	86.04	13.40	2.0	0.01	< 0.01	< 0.01	1	0.87	4.830	2.029	+85	28.6
53-54	2.5	1700		0.12	86.74	13.54	2.0	0.01	< 0.01	< 0.01	<1	1.5	4.858	2.020	+85	28.3

TABLE 55. SECOND-STAGE HYDROTREATING

Product Properties

Run 1

Test Period	7	9	16	18	25	27
API Gravity at 60°F	33.2	33.1	32.1	32.0	33.6	33.8
Sp. Gr. 60/60°F	0.8591	0.8597	0.8649	0.8654	0.8571	0.8560
Distillation (D-1160), °F						
IBP	345	357	345	346	304	310
5%	448	464	475	480	450	440
10%	491	499	509	508	489	479
20%	543	548	553	561	538	530
30%	581	586	592	604	574	569
40%	625	628	636	639	615	609
50%	669	670	682	682	669	657
60%	718	719	732	729	720	709
70%	765	770	788	780	770	760
80%	821	830	850	840	831	828
90%	899	912	939	928	910	910
95%	959	988	1030	1012	993	980
EP	-	-	-	-	1019	1012
% Recovered	95.0	95.0	95.0	95.0	96.0	97.0
% Bottoms	5.0	5.0	5.0	5.0	4.0	3.0
Distillation (GC), °F						
IBP	249	275	266	266	237	240
5%	399	412	398	399	372	378
10%	446	452	446	447	421	426
20%	507	514	519	520	482	488
30%	560	564	569	571	546	548
40%	599	601	605	608	584	586
50%	642	646	652	656	626	627
60%	691	694	703	708	677	676
70%	744	748	761	768	738	736
80%	808	809	820	826	805	801
90%	880	880	882	907	889	879
95%	940	939	947	964	957	945
EP	1062	1060	-	-	-	-

TABLE 56. SECOND-STAGE HYDROTREATING

Product Properties

Run 1 (Continued)

Test Period	34	36	43	45
API Gravity at 60°F	33.3	33.6	32.5	32.6
Sp. Gr. 60/60°F	0.8586	0.8571	0.8628	0.8623
Distillation (D-1160), °F				
IBP	324	340	322	330
5%	450	440	450	453
10%	486	489	492	488
20%	545	549	541	538
30%	590	592	581	580
40%	630	631	622	629
50%	678	670	670	698
60%	725	715	723	730
70%	778	762	781	739
80%	838	821	840	812
90%	910	899	910	930
95%	1000	990	995	1003
EP	-	-	-	-
% Recovered	95.0	95.0	95.0	95.0
% Bottoms	5.0	5.0	5.0	5.0
Distillation (GC), °F				
IBP	228	222	244	240
5%	374	365	383	383
10%	423	421	426	427
20%	497	496	501	503
30%	545	545	555	558
40%	588	589	597	600
50%	631	633	641	647
60%	682	683	691	688
70%	735	738	745	755
80%	797	800	806	815
90%	863	870	871	886
95%	924	932	931	945
EP	1039	1057	1061	-

TABLE 57. SECOND-STAGE HYDROTREATING

Product Properties

Run 2

Test Period	7	9	16	18	25
API Gravity at 60°F	31.8	31.4	30.5	30.1	32.6
Sp. Gr. 60/60°F	0.8665	0.8686	0.8735	0.8756	0.8623
Distillation (D-1160), °F					
IBP	360	308	331	368	321
5%	469	463	469	469	452
10%	503	509	508	510	492
20%	552	560	557	561	542
30%	591	608	594	600	583
40%	637	660	635	644	627
50%	690	710	680	690	669
60%	742	760	729	741	721
70%	800	805	780	797	772
80%	849	858	845	851	830
90%	930	957	930	930	900
95%	93% - 980	92% - 990	94% - 990	94% - 989	94% - 972
EP					
% Recovered	93.0	92.0	94.0	94.0	94.0
% Bottoms	7.0	8.0	6.0	6.0	6.0
Distillation (GC), °F					
IBP	264	232	243	266	233
5%	403	364	382	411	385
10%	438	419	427	460	435
20%	512	494	505	521	499
30%	571	545	564	570	555
40%	611	592	608	613	595
50%	660	637	660	657	643
60%	712	691	714	711	696
70%	768	746	771	768	756
80%	826	807	828	819	816
90%	900	875	904	887	898
95%	952	935	957	939	969
EP	-	1049	-	-	-

TABLE 58. SECOND-STAGE HYDROTREATING

Product Properties

Run 3

Test Period	7	9	20	22
API Gravity at 60°F	33.1	32.6	33.2	33.3
Sp. Gr. 60/60°F	0.8597	0.8623	0.8591	0.8586
Distillation (D-1160), °F				
IBP	340	388	304	326
5%	449	461	451	453
10%	498	500	489	490
20%	549	546	539	542
30%	586	582	578	585
40%	628	627	620	620
50%	670	671	665	661
60%	720	720	712	714
70%	774	770	770	770
80%	838	835	840	825
90%	910	918	940	900
95%	1000	1023	93% - 991	990
EP				
% Recovered	95.0	95.0	93.0	95.0
% Bottoms	5.0	5.0	7.0	5.0
Distillation (GC), °F				
IBP	234	247	211	231
5%	380	396	366	374
10%	431	440	418	426
20%	496	501	484	490
30%	549	556	538	542
40%	596	598	583	589
50%	640	644	629	633
60%	683	695	675	680
70%	752	754	734	739
80%	811	811	798	801
90%	872	872	851	854
95%	933	932	918	918
EP	-	-	-	-

TABLE 59. SECOND-STAGE HYDROTREATING

Product Properties

Run 4

Test Period	7	9	19	21	25
API Gravity at 60°F	31.8	30.0	32.6	32.5	32.7
Sp. Gr. 60/60°F	0.8665	0.8762	0.8623	0.8628	0.8618
Distillation (D-1160), °F					
IBP	350	365	332	314	321
5%	466	462	459	461	438
10%	505	501	490	492	472
20%	547	545	537	526	520
30%	585	585	582	571	557
40%	628	630	628	612	591
50%	673	676	669	650	633
60%	722	722	718	700	688
70%	779	772	769	752	740
80%	840	829	828	820	800
90%	915	899	908	900	868
95%	968	980	1014	972	931
EP	1035	96% - 1003	1031	1000	1000
% Recovered	99.0	96.0	96.0	96.0	98.0
% Bottoms	1.0	4.0	4.0	4.0	2.0
Distillation (GC), °F					
IBP	247	257	227	238	207
5%	420	422	393	391	367
10%	460	462	447	445	428
20%	520	521	508	505	489
30%	573	574	561	556	541
40%	613	612	602	596	582
50%	660	659	647	638	626
60%	712	710	698	687	674
70%	771	766	756	740	728
80%	830	823	816	800	789
90%	924	905	896	867	855
95%	1009	966	961	928	915
EP	-	-	-	-	1048

TABLE 60. SECOND-STAGE HYDROTREATING

Product Properties

Run 5

Test Period	7	9	16	18	27	32
API Gravity at 60°F	32.4	33.0	31.7	31.4	33.9	35.7
Sp. Gr. 60/60°F	0.8633	0.8602	0.8670	0.8686	0.8555	0.8463
Distillation (D-1160), °F						
IBP	321	341	358	376	349	302
5%	455	456	458	481	439	408
10%	491	497	492	510	480	442
20%	545	549	550	569	533	500
30%	588	590	590	610	580	542
40%	631	630	632	651	620	582
50%	674	680	680	715	662	621
60%	721	731	730	749	709	667
70%	775	781	780	796	760	712
80%	840	847	838	850	820	770
90%	920	930	909	933	902	840
95%	1014	1010	1018	1000	1000	895
EP						930
% Recovered	95.0	95.0	95.0	95.0	95.0	97.0
% Bottoms	5.0	5.0	5.0	5.0	5.0	3.0
Distillation (GC), °F						
IBP	257	256	278	291	246	205
5%	409	411	419	423	400	342
10%	453	454	458	464	447	407
20%	511	511	518	521	504	466
30%	564	563	572	575	554	514
40%	602	601	611	615	593	562
50%	646	643	659	663	635	601
60%	695	691	710	714	684	646
70%	748	742	769	772	739	699
80%	805	801	828	829	800	763
90%	866	864	911	910	869	835
95%	923	922	973	966	930	894
EP	1032	1029	-	1137	1047	1025

TABLE 61. SECOND-STAGE HYDROTREATING

Product Properties

Run 5 (Continued)

Test Period	36	38	42
API Gravity at 60°F	39.5	39.2	38.8
Sp. Gr. 60/60°F	0.8275	0.8289	0.8309
Distillation (D-1160), °F			
IBP	195	267	238
5%	342	350	346
10%	396	400	391
20%	450	452	450
30%	498	490	490
40%	540	530	527
50%	581	574	561
60%	622	628	603
70%	670	681	652
80%	735	748	721
90%	816	820	810
95%	871	871	866
EP	888	908	905
% Recovered	96.0	98.0	98.0
% Bottoms	4.0	2.0	2.0
Distillation (GC), °F			
IBP	191	191	203
5%	293	297	302
10%	355	357	359
20%	423	424	429
30%	466	470	476
40%	509	510	514
50%	551	550	558
60%	587	585	593
70%	632	628	639
80%	693	683	698
90%	771	758	780
95%	818	807	826
EP	927	913	933

TABLE 62. SECOND-STAGE HYDROTREATING

Product Properties

Run 6

Test Period		7	9	15
API Gravity at 60°F		30.6	31.8	33.2
Sp. Gr. 60/60°F		0.8729	0.8665	0.8591
Distillation (D-1160), °F				
IBP		407	392	325
5%		490	441	440
10%		520	482	479
20%		559	544	532
30%		599	595	578
40%		642	639	620
50%		688	679	663
60%		736	724	714
70%		790	772	761
80%		857	839	820
90%		961	912	890
95%	92% -	1007	1001	946
EP		-		1010
% Recovered		92.0	95.0	97.0
% Bottoms		8.0	5.0	3.0
Distillation (GC), °F				
IBP		262	309	270
5%		409	422	410
10%		452	459	458
20%		506	512	512
30%		562	565	557
40%		600	601	594
50%		645	643	634
60%		695	689	682
70%		749	740	733
80%		806	796	791
90%		871	852	854
95%		932	901	909
EP		-	971	1021

TABLE 63. SECOND-STAGE HYDROTREATING

Product Properties

Run 7

Test Period	7	9	14
API Gravity at 60°F	32.5	32.5	33.6
Sp. Gr. 60/60°F	0.8628	0.8628	0.8571
Distillation (D-1160), °F			
IBP	355	310	315
5%	451	445	430
10%	491	491	475
20%	540	541	529
30%	580	581	570
40%	619	625	610
50%	668	671	649
60%	711	715	700
70%	760	765	747
80%	821	827	803
90%	902	908	880
95%	981	1000	945
EP			1000
% Recovered	95.0	95.0	97.0
% Bottoms	5.0	5.0	3.0
Distillation (GC), °F			
IBP	177	257	246
5%	364	415	396
10%	435	460	451
20%	502	515	509
30%	556	564	557
40%	604	602	596
50%	658	644	640
60%	718	693	690
70%	786	748	746
80%	854	808	809
90%	952	875	880
95%	1030	935	946
EP	-	1041	-

TABLE 64. SECOND-STAGE HYDROTREATING

Product Properties

Run 8

Test Period	7
API Gravity at 60°F	33.0
Sp. Gr. 60/60°F	0.8602
Distillation (D-1160), °F	
IBP	346
5%	460
10%	496
20%	540
30%	580
40%	618
50%	655
60%	699
70%	748
80%	810
90%	898
95%	980
EP	1010
% Recovered	96.0
% Bottoms	4.0
Distillation (GC), °F	
IBP	257
5%	414
10%	460
20%	516
30%	563
40%	601
50%	639
60%	690
70%	745
80%	810
90%	884
95%	947
EP	1116

TABLE 65. SECOND-STAGE HYDROTREATING

Product Properties

Run 9

Test Period	7	9	16	18	25	27
API Gravity at 60°F	32.6	32.5	33.8	33.4		
Sp. Gr. 60/60°F	0.8623	0.8628	0.8560	0.8581		
Distillation (D-1160), °F						
IBP	319	332	324	316		
5%	437	441	420	428		
10%	478	481	455	472		
20%	541	546	513	534		
30%	597	596	570	582		
40%	642	647	621	635		
50%	689	698	670	683		
60%	739	751	719	728		
70%	799	805	769	775		
80%	851	865	829	834		
90%	920	940	902	910		
95%	985	985	97% - 950	991		
EP	-	-	983	-		
% Recovered	95.0	95.0	97.0	95.0		
% Bottoms	5.0	5.0	3.0	5.0		
Distillation (GC), °F						
IBP	241	282	238	347	267	339
5%	385	396	370	421	383	417
10%	444	439	426	455	427	452
20%	515	497	493	506	488	504
30%	578	551	553	553	541	551
40%	637	594	601	592	585	591
50%	688	638	653	632	631	633
60%	739	684	711	676	681	680
70%	794	732	771	725	736	732
80%	852	788	832	782	799	789
90%	937	857	908	844	871	851
95%	1008	909	962	885	926	894
EP	1176	1025	-	962	1055	969

TABLE 66. SECOND-STAGE HYDROTREATING

Product Properties

Run 9 (Continued)

Test Period	34	36	43	45
API Gravity at 60°F	35.1	35.1	37.3	37.1
Sp. Gr. 60/60°F	0.8493	0.8493	0.8383	0.8393
Distillation (D-1160), °F				
IBP	305	310		
5%	371	379		
10%	421	427		
20%	487	492		
30%	539	546		
40%	582	588		
50%	630	631		
60%	678	679		
70%	729	731		
80%	785	790		
90%	854	858		
95%	892	899		
EP	949	948		
% Recovered	99.0	99.0		
% Bottoms	1.0	1.0		
Distillation (GC), °F				
IBP	226	259	214	268
5%	325	356	326	363
10%	374	404	379	408
20%	443	462	446	464
30%	491	509	493	511
40%	546	553	543	558
50%	593	591	583	598
60%	644	631	629	645
70%	698	676	680	697
80%	764	727	741	760
90%	835	787	815	828
95%	885	825	866	870
EP	983	921	979	949

TABLE 67. SECOND-STAGE HYDROTREATING

Product Properties

Run 10

Test Period	7	9	16	18	25	27
API Gravity at 60°F	32.9	32.4	33.3	33.8	35.2	35.7
Sp. Gr. 60/60°F	0.8607	0.8633	0.8596	0.8560	0.8488	0.8463
Distillation (D-1160), °F						
IBP	253	280	245	166		
5%	415	410	390	399		
10%	459	464	446	435		
20%	528	531	507	500		
30%	580	586	560	559		
40%	629	638	610	603		
50%	677	688	660	658		
60%	729	738	715	710		
70%	781	789	770	760		
80%	841	849	825	820		
90%		921	862	899		
95%	955	970	942	949		
EP	988	1011	1020	998		
% Recovered	98.0	98.0	99.0	98.0		
% Bottoms	2.0	2.0	1.0	2.0		
Distillation (GC), °F						
IBP	272	257	252	259	259	297
5%	404	386	389	352	378	411
10%	455	434	439	405	427	449
20%	521	495	505	471	490	499
30%	577	546	561	522	542	547
40%	633	590	607	576	588	583
50%	696	633	660	628	635	624
60%	761	685	716	685	692	668
70%	819	738	776	744	754	715
80%	870	798	836	808	821	769
90%	941	868	909	880	901	828
95%	1002	938	960	932	964	869
EP	-	-	-	1039	-	950

TABLE 68. SECOND-STAGE HYDROTREATING

Product Properties

Run 10 (Continued)

Test Period	34	36	42-43	44-45	51-52	53-54
API Gravity at 60°F	35.3	34.8	37.1	37.3	34.1	34.3
Sp. Gr. 60/60°F	0.8483	0.8509	0.8393	0.8383	0.8545	0.8534
Distillation (D-1160), °F						
IBP	223	236	242	280	230	256
5%	330	362	340	356	378	389
10%	386	418	402	398	429	430
20%	460	480	462	458	494	497
30%	519	529	509	510	549	550
40%	570	575	549	550	592	595
50%	615	625	594	598	640	641
60%	668	679	640	642	690	690
70%	720	729	691	690	740	739
80%	779	788	748	749	798	799
90%	850	859	819	821	858	865
95%	899	910	861	868	895	906
EP	971	978	932	931	969	975
% Recovered	99.0	99.0	99.0	99.0	99.0	99.0
% Bottoms	1.0	1.0	1.0	1.0	1.0	1.0
Distillation (GC), °F						
IBP	205	242	212	209	203	223
5%	335	358	328	321	359	335
10%	394	408	383	375	417	383
20%	468	468	452	443	485	453
30%	520	515	500	489	544	507
40%	571	561	544	531	596	561
50%	619	601	587	572	646	608
60%	674	649	633	611	696	665
70%	738	704	692	657	750	724
80%	808	768	762	715	809	793
90%	885	842	840	794	882	873
95%	943	900	907	878	945	935
EP	-	-	-	-	1092	-

TABLE 69. SECOND-STAGE HYDROTREATING

Summary of Product Properties

Source	Occidental Part 1	Occidental Part 2	Paraho Part 3
API Gravity at 60°F	34.3	31.8	33.3
Specific Gravity at 60°F	0.8534	0.8664	0.8588
Distillation, °F			
IBP	255	245	250
5%	380	405	383
10%	430	449	432
30%	541	561	556
50%	619	646	654
70%	723	752	766
90%	858	883	900
95%	919	940	956
EP	1052	1079	1085
Pour Point, °F	+75	+80	+85
Viscosities			
Kinematic at 122°F, cst	6.327	7.757	6.095
Kinematic at 210°F, cst	2.341	2.766	2.433
Universal at 122°F, sus	46.7	51.4	46.0
Universal at 210°F, sus	34.0	35.4	34.3
Aromatics, vol-%	20.4	40.1	27.2
Paraffins and Naphthenes, vol-%	79.6	59.9	72.8
Carbon, wt-%	86.0	86.4	86.9
Hydrogen, wt-%	13.9	13.6	13.1
Nitrogen, ppm	100	900	700
Sulfur, ppm	137	500	< 300
Oxygen, ppm	144	651	-
Heptane Insolubles, wt-%	0.03	0.06	0.06
Conradson Carbon, wt-%	< 0.01	< 0.01	< 0.01
Ash, wt-%	< 0.01	0.001	0.002
Bromine No.	1.0	1.1	1.0
Metals by Emission			
Fe, ppm	0.76	0.48	< 0.05
Ni, ppm	0.05	0.05	< 0.05
V, ppm	0.04	< 0.04	< 0.05
Pb, ppm	0.10	< 0.1	< 0.1
Cu, ppm	0.42	0.29	< 0.05
Na, ppm	0.36	< 0.1	< 0.1
Mo, ppm	0.10	< 0.1	< 0.1
Arsenic, ppm	< 1	< 1	< 1

TABLE 70. HIGH PRESSURE HYDROTREATING -- PRODUCT YIELDS

Catalyst: DCA

Feedstock: Low Pressure Hydrotreated Occidental Shale Oil

Product Nitrogen: 900 ppm

<u>Yields, Wt-% of Feed</u>

H_2O	0.22
NH_3	1.21
H_2S	0.03
C_1	0.17
C_2	0.23
C_3	0.21
C_4	0.29
C_5	0.63
C_6+	<u>98.36</u>
Total	101.35

Chemical Hydrogen Consumption = 798 SCFB

TABLE 71. HIGH PRESSURE HYDROTREATING -- PRODUCT YIELDS

Catalyst: DCA

Feedstock: Low Pressure Paraho Shale Oil

Product Nitrogen: 700 ppm

Yields, Wt-% of Feed

NH_3	2.23
H_2S	0.05
C_1	0.16
C_2	0.23
C_3	0.21
C_4	0.28
C_5	0.62
C_6+	97.81
Total	101.59

Chemical Hydrogen Consumption = 960 SCFB

TABLE 72. SECOND-STAGE HYDROTREATING

Used Catalyst Analysis

	Upper Section	Upper Mid.	Mid. Section	Lower Mid.	Lower Section
Arsenic, ppm	220	112	74	65	38
Carbon, wt-%	6.84	7.11	5.89	7.00	6.03
Iron, wt-%	0.53	0.17	0.17	0.17	0.22
Nitrogen, wt-%	1.73	1.11	0.63	1.20	1.17
Sulfur, wt-%	6.22	6.12	6.43	7.32	6.58

TABLE 73. COMMERCIAL DESIGN OPERATING CONDITIONS

Parallel Flow Hydrocracking

Separator Pressure, psi $(P-P_B)$	-150
Liquid Hourly Space Velocity (L/L_B)	1.4
Combined Feed Ratio (C/C_B)	1.0
Hydrogen Circulation (H/H_B)	1.3
Jet Fuel Product End Point, °F	550
Catalyst Type	UOP-DCC

Reactor Temperature, °F $(T-T_B)$: Adjusted to achieve 100% conversion to 550°F

TABLE 74. HIGH PRESSURE HYDROTREATED SHALE OIL INSPECTIONS

Hydrocracker Feedstock

Feedstock	Occidental	Paraho
API Gravity at 60°F	33.1	33.3
Specific Gravity at 60°F		0.8588
Distillation, °F		
IBP	320	250
50%	655	654
EP	938	∿ 1085
Pour Point, °F	+80	+85
Bromine Number	1.3	1.0
Sulfur, wt-%	< 0.03	< 0.03
Nitrogen, wt-%	0.067	0.07
Arsenic, ppm	< 1	< 1
Conradson Carbon, wt-%	0.03	< 0.01
Total Oxygen, ppm	187	410

TABLE 75. JET FUEL ANALYSES

	Occidental 700 JP-8 3794-50C	Paraho 700 JP-8	Occidental 700 JP-4 3794-53	Occidental 700 JP-4(a) 3794-90	Occidental(b) 1300 JP-8(c) 3794-52	Occidental(b) 1300 JP-4(d) 3794-91	Occidental 900 JP-8 3794-54C	Occidental 900 JP-8 3794-55C	U.S. Military Specs JP-4	U.S. Military Specs JP-8
Feedstock										
HP Hydrotreated Shale Oil										
Nitrogen Content in Feed, ppm	700	700	700	700	1300	1300	900	900		
Jet Fuel Type	JP-8	JP-8	JP-4	JP-4(a)	JP-8(c)	JP-4(d)	JP-8	JP-8	JP-4	JP-8
UOP Identification	3794-50C		3794-53	3794-90	3794-52	3794-91	3794-54C	3794-55C		
Analyses										
API at 60°F	45.1	45.6	52.5	49.3	44.9	55.3	43.4	43.8	45-57	37-51
Sp. Gr. 60/60°F	0.8012	0.7990	0.7690	0.7826	0.8022	0.7575	0.8090	0.8072	0.751-0.802	0.775-0.840
Distillation, °F										
IBP	295	300	157	188	297	166	288	304	Report	Report
5%	314	318	190	216	316	180	314	323		
10%	330	334	208	242	326	197	328	336	Report	401 max.
20%	359	362	236	290	334	225	358	365	293 max.	Report
30%	387	384	280	336	385	252	395	394		
40%	412	410	326	374	415	284	427	423		
50%	438	435	365	416	438	314	451	448	374 max.	Report
60%	458	456	396	444	458	352	476	470		
70%	474	474	422	466	476	390	496	490		
80%	492	492	449	486	492	425	517	509	473 max.	Report
90%	512	511	471	508	512	462	538	530		
95%	524	524	486	522	522	482	554	542		
EP	532	555	498	534	533	494	561	550	518 max.	572 max.
% Over	99.0	98.5	99.0	99.0	99.0	99.0	98.5	98.5	0.4 max.*	0.4 max.*
Sulfur, ppm	282	219	111	10.2	207	8.7	129	125		
Nitrogen, ppm	1.1	0.54	0.68	0.3	64.8	0.2	0.4	2.5		
Carbon, wt-%	84.33	85.81	84.47	85.76	85.0	85.46	85.56	85.54		
Hydrogen, wt-%	13.93	14.36	14.27	14.69	13.8	14.75	13.82	13.82	13.6 min.	13.6 min.
Flash Point, °F	100	+98	-	-	96	-	98	104	100 min.	100 min.
Freeze Point, °F	-58	-49	Below -75	-54	-57	< -75	< -56	-45	-72 max.	-58 max.
Smoke Point, mm	25.8	26.5	31.1	29.0	25	33.1	20	20	20 min.	25 min.
Viscosity @-20°C										
cst	5.39	7.473	-	-	5.58	-	6.14	5.88	-	8.0 max.
sus	43.3	49.9	-	-	45.9	-	45.7	44.9		
Naphthalenes, wt-%	0.46	0.5	0.27	0.26	0.41	0.20	2.3	1.7	-	3.0 max.**
Cu Strip Corrosion	1A	4A	1A	1A	1A	1A	1A	1A	1B max.	1B max.
Acid No. mg KOH/g	0.022	-	< 0.001	0.004	0.018	0.031	0.009	0.001	0.015 max.	0.015 max.
Thermal Combustion, Btu/lb	19,570	19,640	20,025	19,980	19,895	20,100	19,762	19,785	18,400 min.	18,400 min.
FIA, vol-%										
A	10.8	11.6	7.2	7.3	13.8	6.5	20.3	18.3	25.0 max.	25.0 max.
O	-	-	0.7	0.5	-	0.8	-	-		
P + N	89.2	88.4	92.1	92.2	86.2	92.7	79.7	81.7		
JFTOT Evaluation										
TDR Spun	4.5									
ASTM Code	2.0									
Temp, °C	260.0									
Min	150.0									
P, MM Hg	0.0									

* Indicates wt-%.
** Indicates vol-%.

(a) Produced at JP-8 mode of operation. (b) 550°F plus fraction.
(c) A blend of JP-8 from 550°F minus and JP-8 from hydrocracking the 550°F plus fraction. (d) Produced at JP-4 mode of operation.

TABLE 76. HYDROCRACKER CHARGE STOCK INSPECTION

Commercial Design Verification Run

SOURCE: High Pressure Hydrotreated Occidental Shale Oil

Identification	550°F+ Fraction	550°F- Fraction
API Gravity at 60°F	29.9	39.4
Sp. Gr. at 60°F	0.8767	0.8280
Distillation (D-1160), °F		
IBP	580	340
5%	620	383
10%	649	408
20%	689	430
30%	713	450
40%	740	458
50%	771	467
60%	800	476
70%	838	484
80%	880	492
90%	960	503
EP	1020	511
% Over	92.0	98.5
Sulfur, wt-ppm	500	10.5
Nitrogen, wt-ppm	1300	215
Carbon, wt-%	86.70	85.88
Hydrogen, wt-%	13.27	13.67
Oxygen, wt-ppm	170	215
Bromine Number	-	
Basic Nitrogen, wt-ppm	370	
Arsenic, wt-ppm	1.3	
Iron, wt-ppm	5.1	
C_5 Insolubles, wt-%	0.04	
C_7 Insolubles, wt-%	< 0.01	
Conradson Carbon, wt-%	0.12	
Pour Point, °F	+75	
Viscosity at 122°F		
cst	16.38	
sus	83.0	
Viscosity at 210°F		
cst	4.54	
sus	41.2	
Thermal Combustion, Btu/lb	-	19,710
Freeze Point, °F	-	-42
FIA, vol-%		
Aromatics	-	19.5
Paraffins + Naphthenes	-	80.5
Naphthalenes, wt-%	-	0.6
Cu Strip Corrosion, 3 hrs at 122°F		1A
Smoke Point		18.9

TABLE 77. NAPHTHA ANALYSES

Feedstock: HP Hydrotreated Shale Oil ($\sim$ 700 ppm Nitrogen)

Operating Mode: Max JP-8

Feedstock Identification	Occidental	Occidental 550°F+ Fraction*	Paraho
C5/C6			
API at 60°F	78.1	80.5	79.8
Sp. Gr. at 60/60°F	0.6751	0.6674	0.6697
Sulfur, ppm	5.95	4.29	2.3
Nitrogen, ppm	1.13	0.21	0.8
RON, Clear	71.2	72.0	
RON, Leaded + 3 mL	87.6	90.5	
C7-250°F			
API at 60°F	63.4	64.6	63.5
Sp. Gr. at 60/60°F	0.7260	0.7216	0.7256
Distillation, °F			
IBP	199	196	198
50%	209	210	204
EP	258	263	265
Sulfur, ppm	50	16.8	102
Nitrogen, ppm	0.2	0.24	0.25
FIA, vol-%			
A	2.1	-	5.2
O	trace	-	0.6
P + N	97.9	-	94.2
RON, Clear	60.8	58.1	-
RON, Leaded + 3 mL	81.2	79.0	-

* $\sim$ 1300 ppm nitrogen

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TABLE 78. PRODUCT DISTRIBUTIONS

Feedstock Identification	Hydrotreated Occidental Shale Oil				Hydrotreated Paraho Shale Oil
	∿ 900 ppm N	∿ 700 ppm N	∿ 700 ppm N	550°F+ fraction	∿ 700 ppm N
Product Desired	JP-8	JP-8	JP-4	JP-8	JP-8
Wt-%					
Fresh Feed	100.00	100.00	100.00	100.00	100.00
H_2 Consumption, wt-%	1.60	1.61	1.80	1.99	1.59
H_2 Consumption, SCFB	902	921	1029	1153	902
NH_3	0.12	0.08	0.08	0.16	0.14
H_2S	0.01	0.01	0.01	-	-
C_1-C_3	2.66	1.50	1.88	2.12	1.87
C_4	4.21	3.73	4.80	6.98	4.39
C_5-C_6	8.55	9.29	11.03	13.07	9.73
C_7-250	9.17*	11.75	12.18	13.26	9.89
250-EP	76.88*	75.25**	71.82***	66.40**	75.57**
Total	101.60	101.61	101.80	101.99	101.59
Vol-%					
Fresh Feed	100.00	100.00	100.00	100.00	100.00
C_4	6.40	5.78	7.43	10.96	6.73
C_5-C_6	11.44	12.62	14.75	18.02	13.12
C_7-250	10.55*	13.76	14.31	15.45	11.30
250-EP	81.00*	80.40**	78.68***	71.87**	80.12**
Total	109.39	112.56	115.17	116.30	111.27

 * C_7-300, 300-EP
 ** Product EP ∿ 550°F
*** Product EP ∿ 520°F

CHARGE TANK · PRE HTR. · 1ST STAGE · 2ND STAGE · COOLER · DEASHED, DEWATERED SHALE OIL
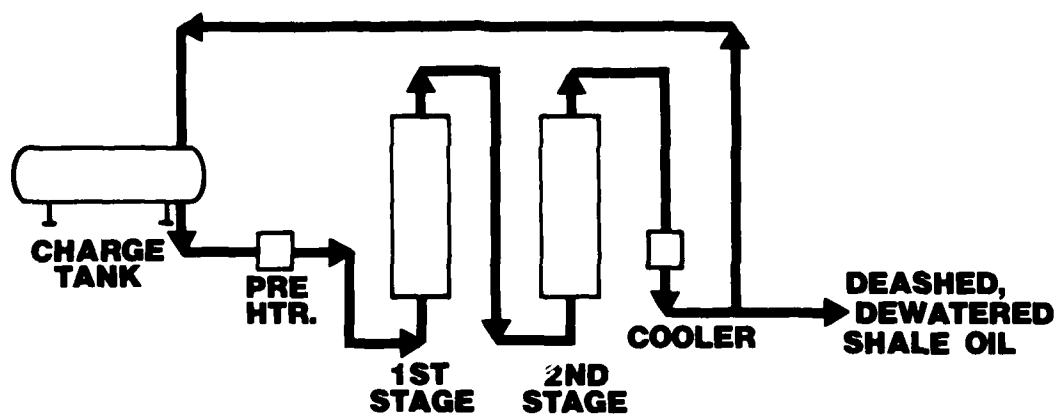
UOP 663-25

FIGURE 5

PILOT PLANT SCHEMATIC FLOW DIAGRAM
FEED PREPARATION

FRESH HYDROGEN →

CHARGE STOCK →

REACTOR

→ OFF GAS

FLASH DRUM

→ LIQUID PRODUCT
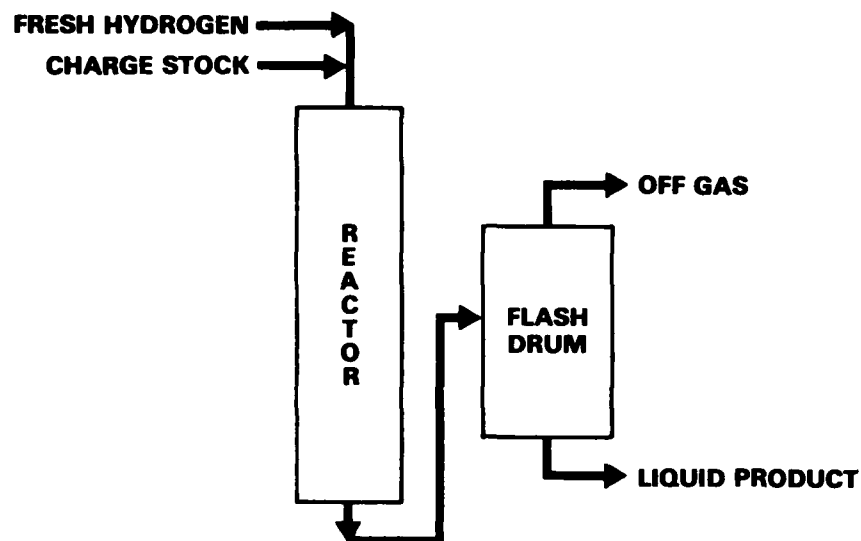
UOP 525-5

FIGURE 6

PILOT PLANT SCHEMATIC FLOW DIAGRAM

LOW PRESSURE HYDROTREATING
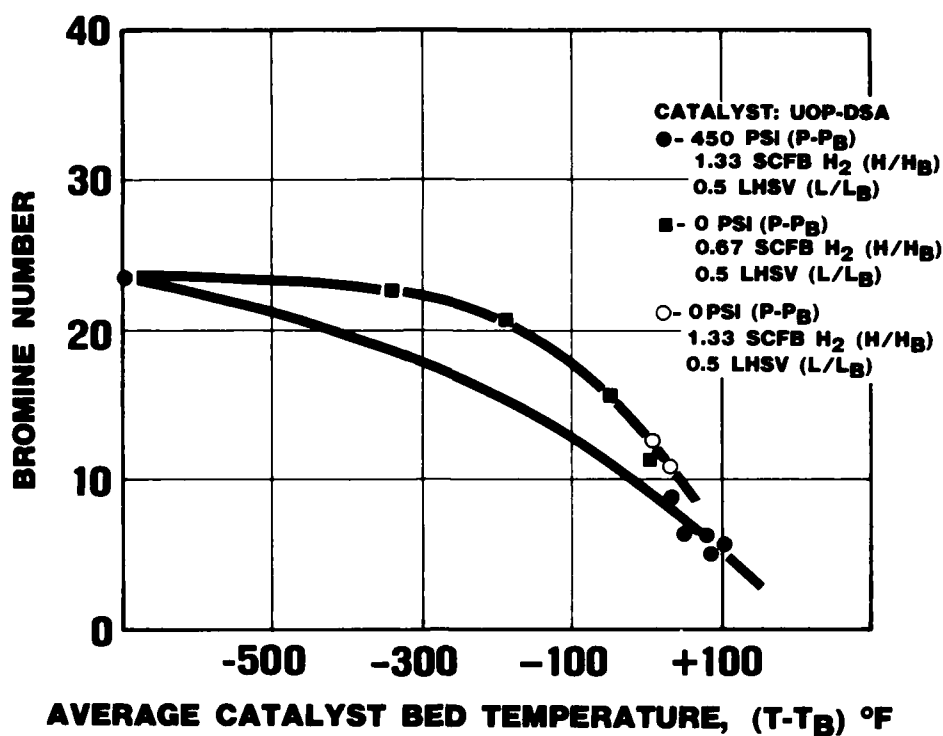
FIGURE 7

FIRST-STAGE HYDROTREATING
EFFECT OF TEMPERATURE AND
HYDROGENATION ON SATURATION
FEED: OCCIDENTAL SHALE OIL
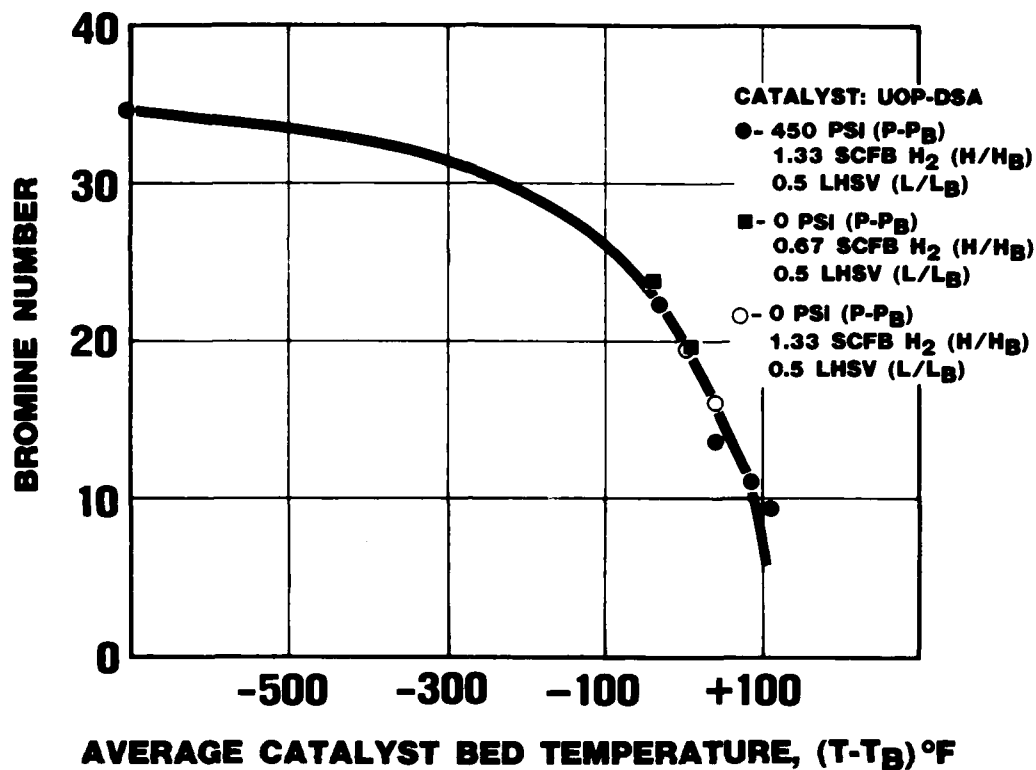
UOP 663-15

FIGURE 8

FIRST-STAGE HYDROTREATING
EFFECT OF TEMPERATURE AND
HYDROGENATION ON SATURATION
FEED: PARAHO SHALE OIL

CATALYST: UOP-DSA

0 PSI ($P-P_B$)

1.33 SCFB H_2 (H/H_B)

+40°F ($T-T_B$)
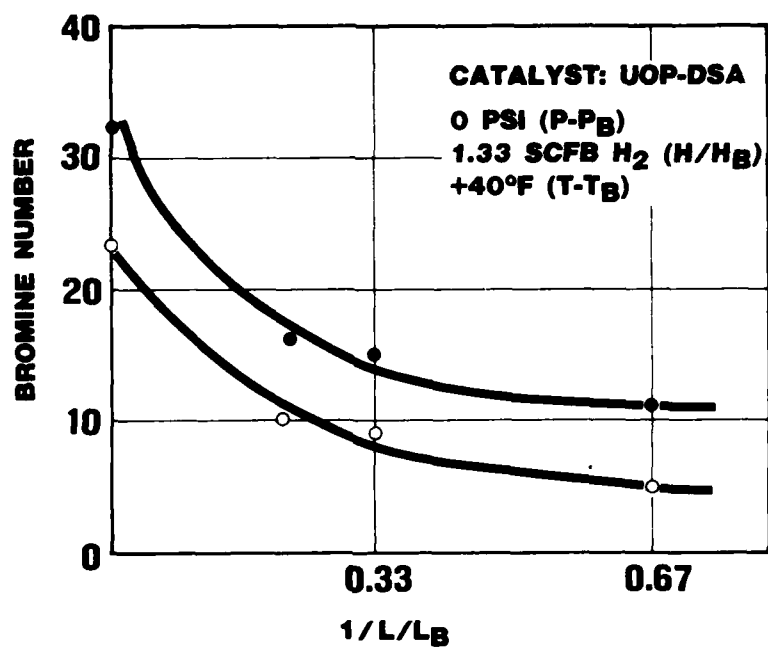
UOP 663-18

FIGURE 9

FIRST-STAGE HYDROTREATING
EFFECT OF LHSV ON SATURATION
FEED: OCCIDENTAL (○) AND PARAHO (●) SHALE OIL

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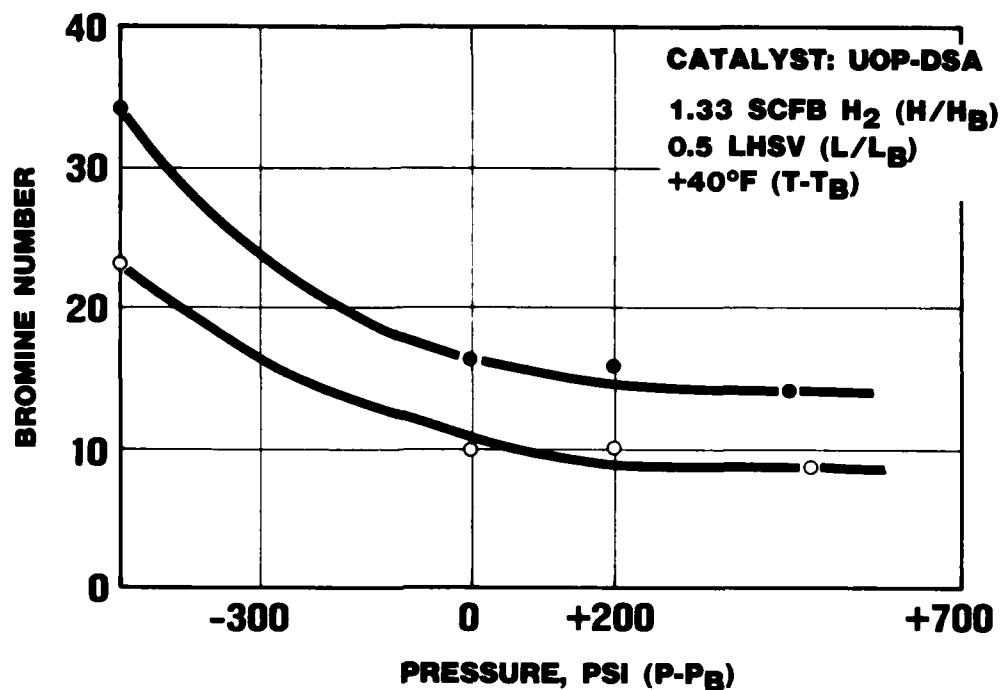
FIGURE 10

FIRST-STAGE HYDROTREATING
EFFECT OF PRESSURE ON SATURATION
FEED: OCCIDENTAL (o) AND PARAHO (•) SHALE OIL

CATALYST: UOP-DRA

O PSI (P-P$_B$)
0.5 LHSV (L/L$_B$)
● 1.33 SCFB H$_2$ (H/H$_B$)

AVERAGE CATALYST BED TEMPERATURE, (T-T$_B$) °F
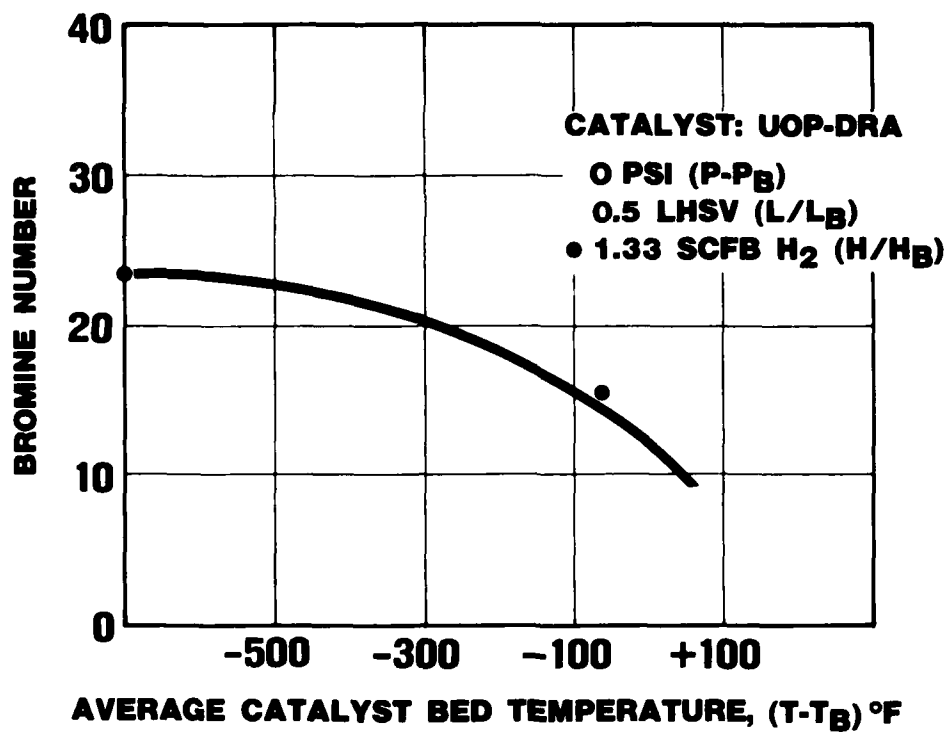
UOP 663-16

FIGURE 11

FIRST-STAGE HYDROTREATING
EFFECT OF TEMPERATURE AND
HYDROGENATION ON SATURATION
FEED: OCCIDENTAL SHALE OIL
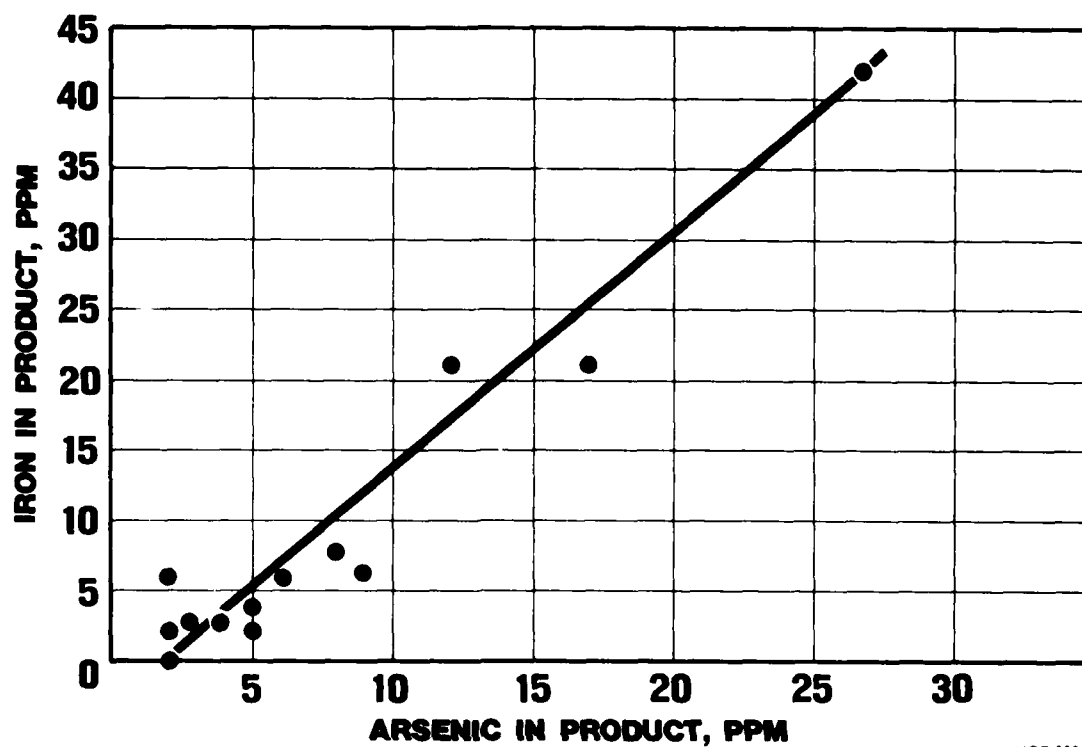
FIGURE 12

FIRST-STAGE HYDROTREATING
PRODUCT IRON vs ARSENIC
OCCIDENTAL SHALE OIL
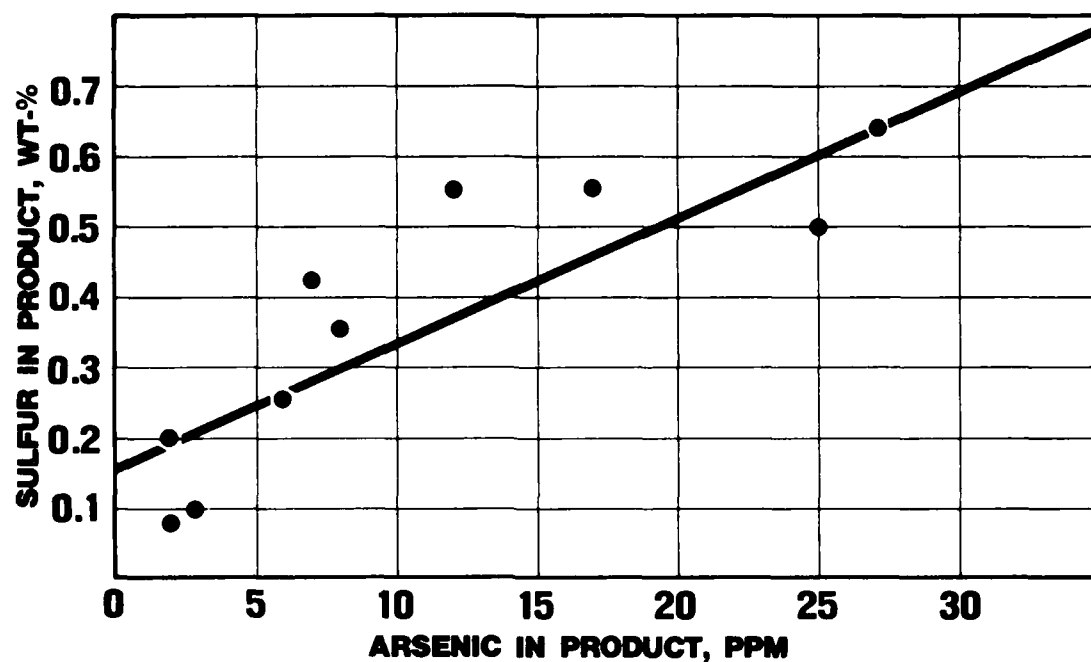
UOP 663-37

FIGURE 13

FIRST-STAGE HYDROTREATING
PRODUCT SULFUR vs ARSENIC
OCCIDENTAL SHALE OIL
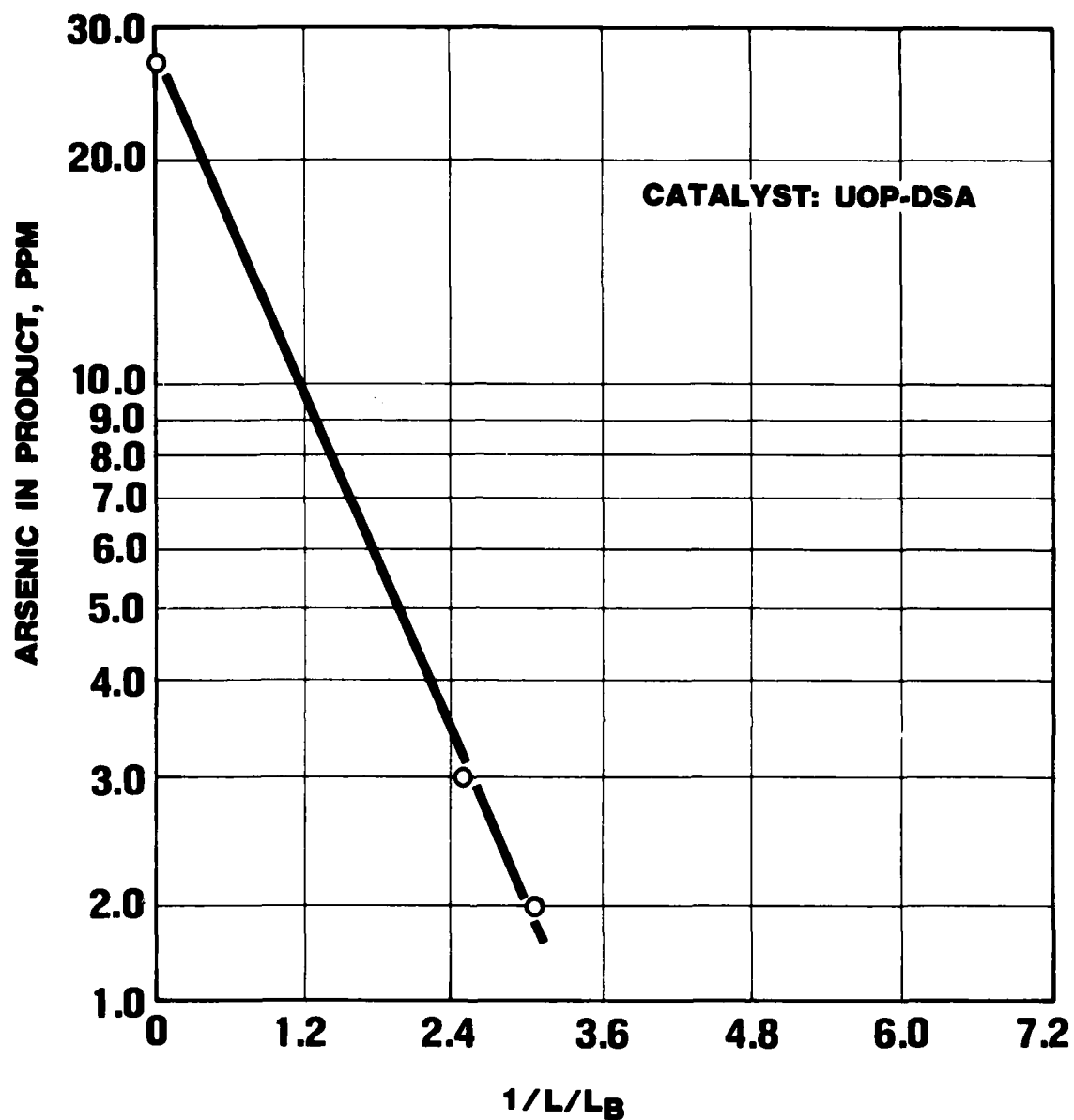
FIGURE 14

FIRST-STAGE HYDROTREATING
ARSENIC vs 1/LHSV

FEED: OCCIDENTAL SHALE OIL
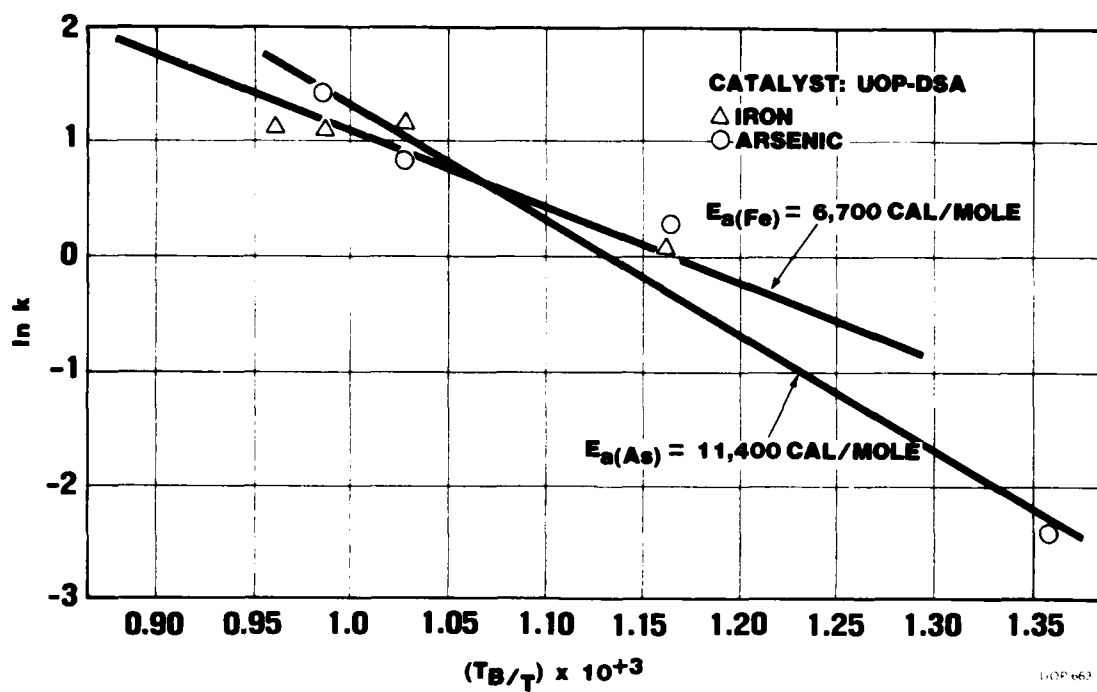
FIGURE 15

FIRST STAGE HYDROTREATING ARRHENIUS PLOT FOR ARSENIC AND IRON REMOVAL
FEED: OCCIDENTAL SHALE OIL
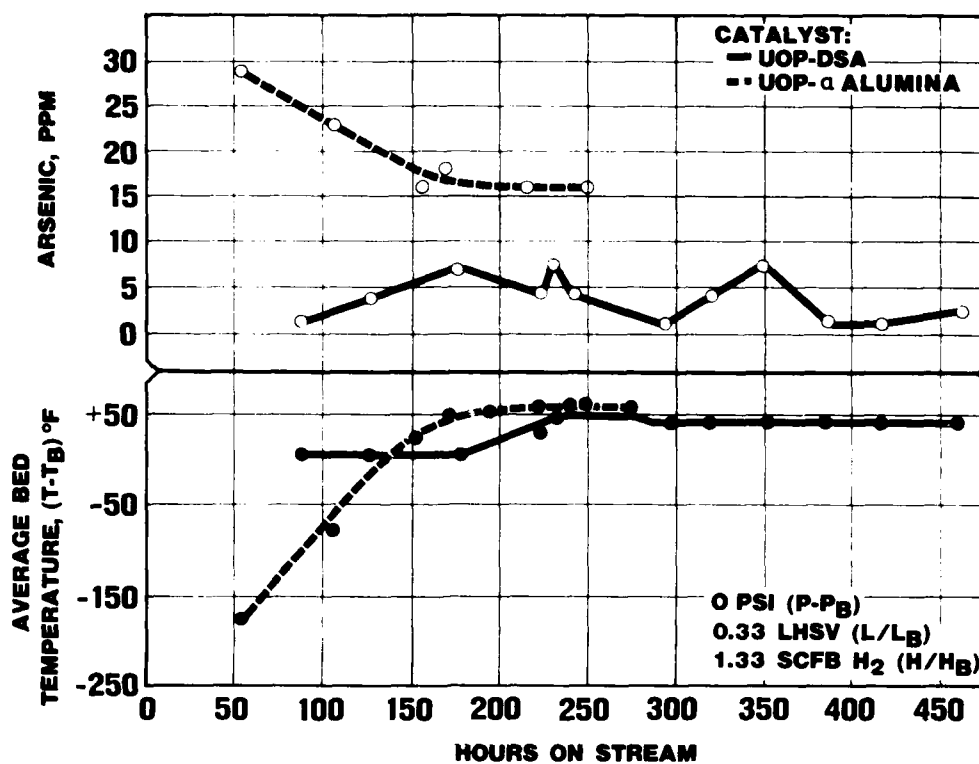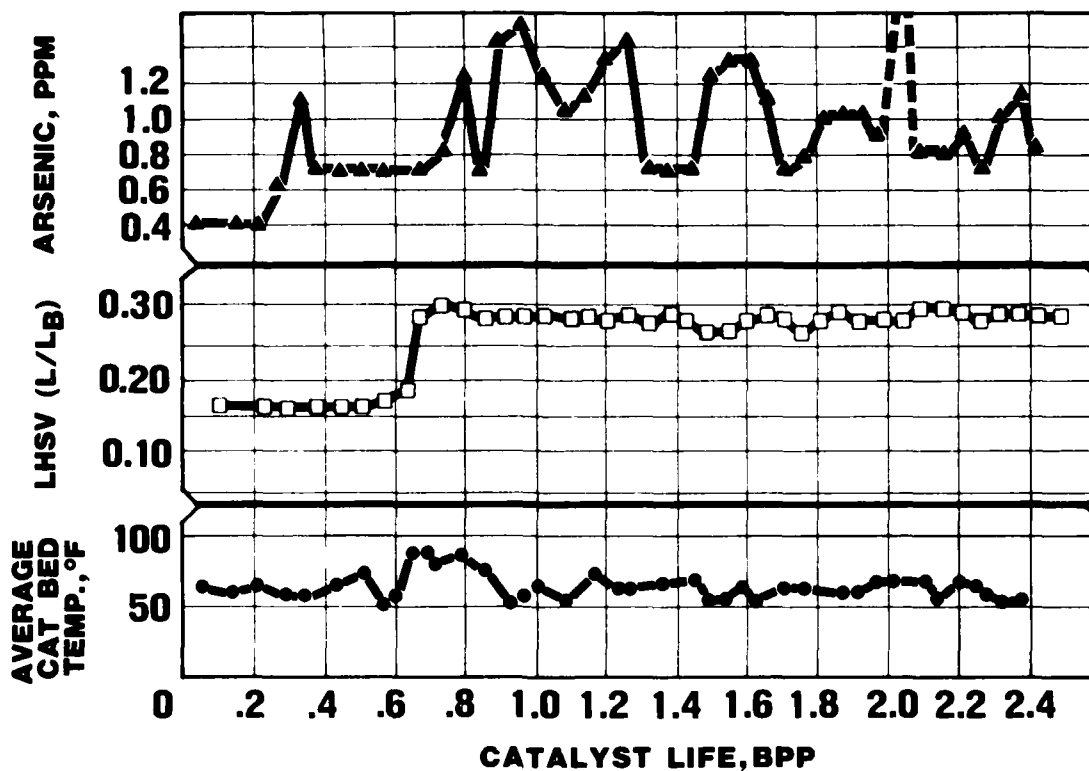
FIGURE 16

FIRST-STAGE HYDROTREATING
FEED: OCCIDENTAL SHALE OIL
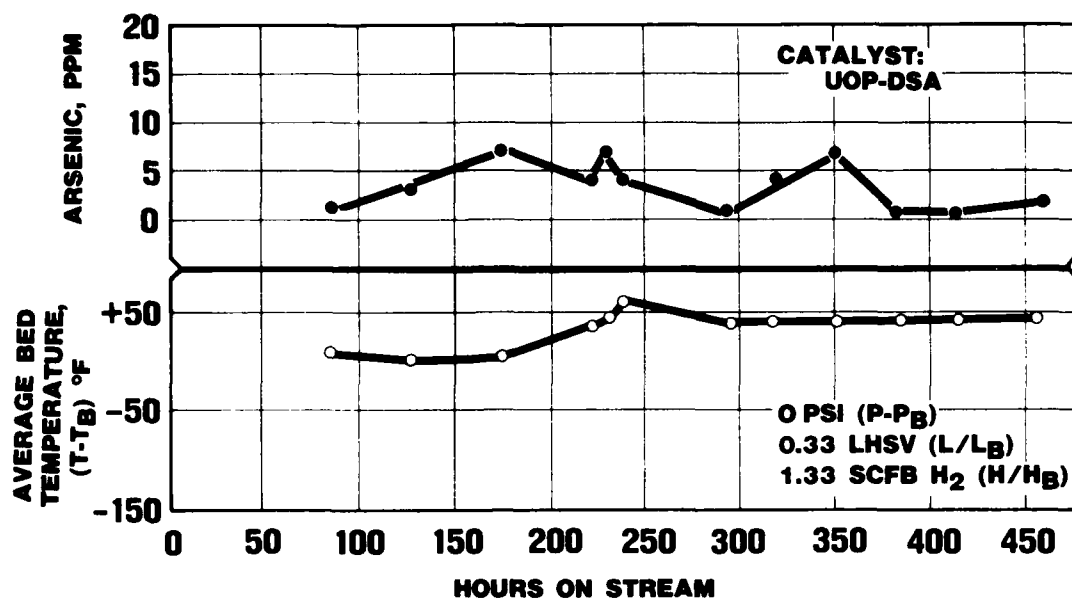
FIGURE 17
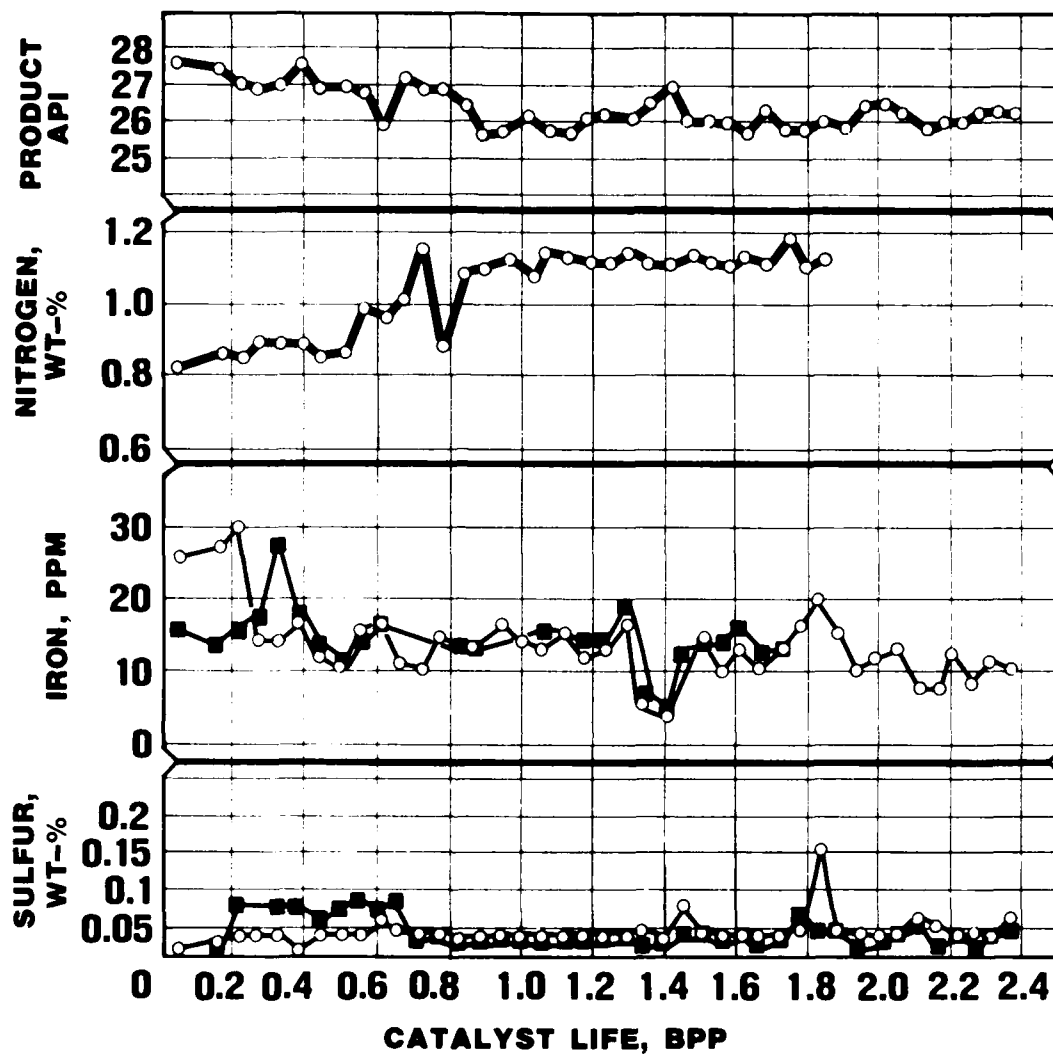
FIRST-STAGE HYDROTREATING
FEED: OCCIDENTAL SHALE OIL
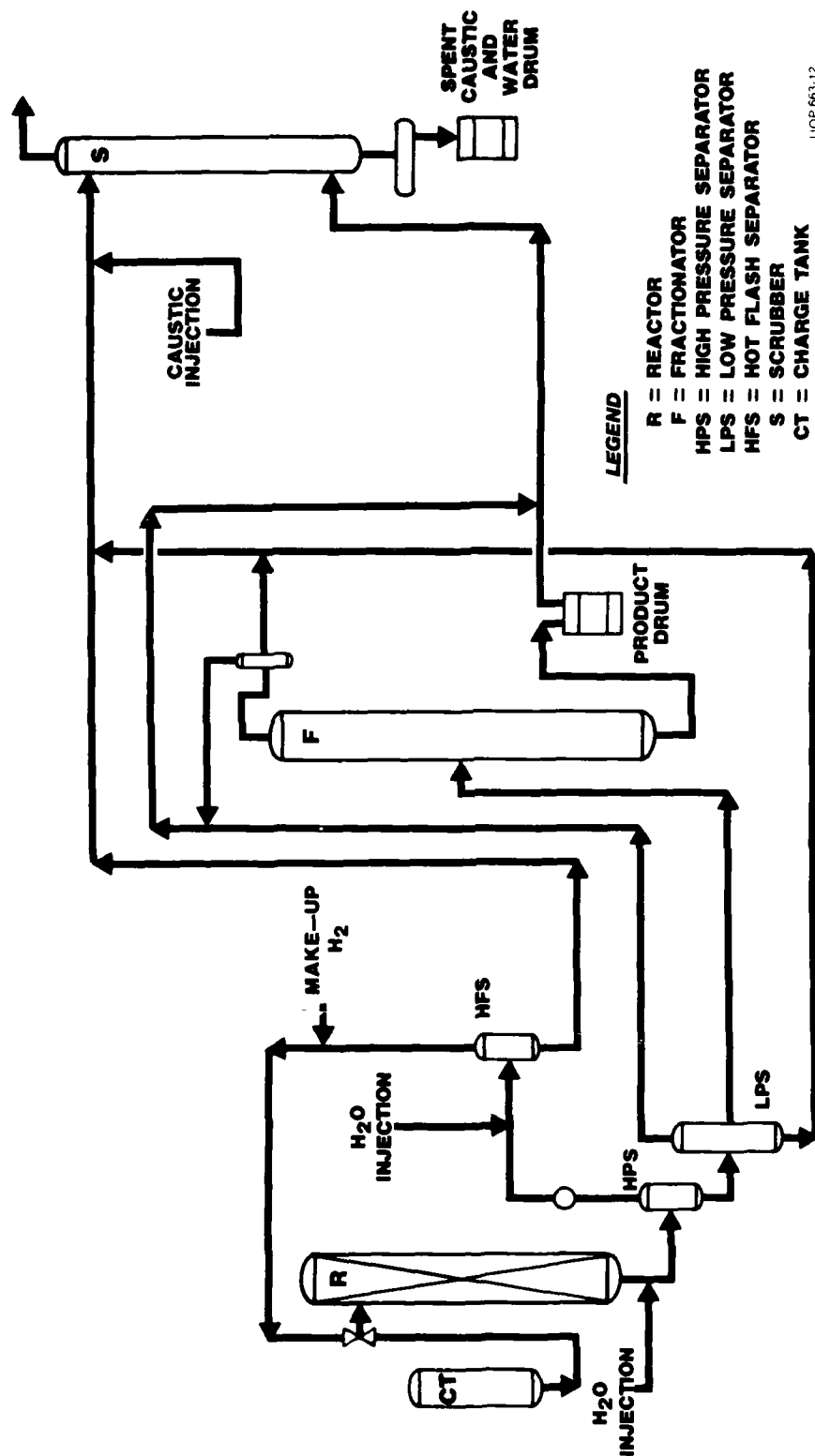
FIGURE 18

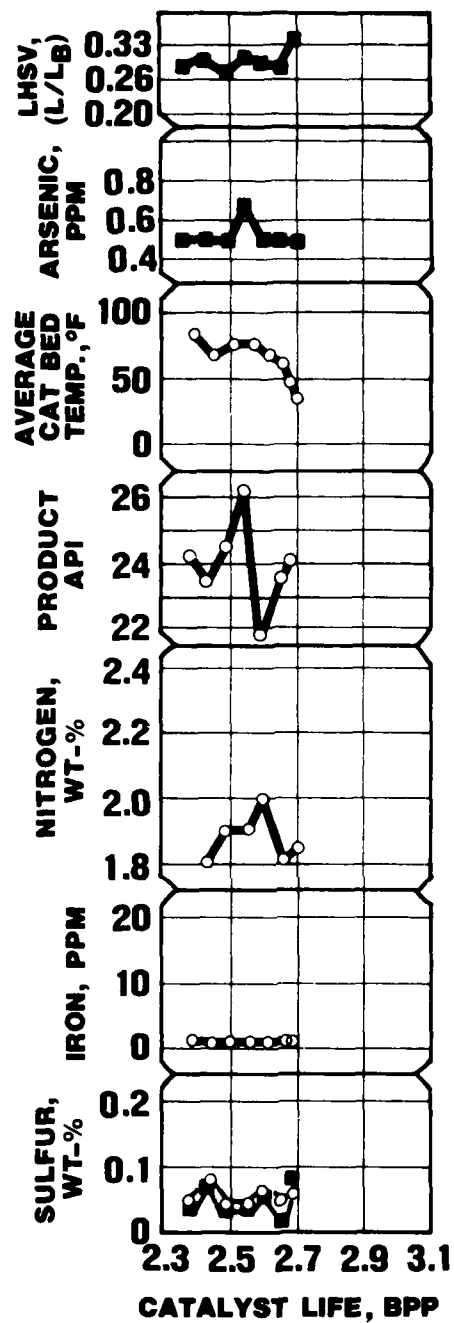# PILOT PLANT SCHEMATIC FLOW DIAGRAM
FIRST STAGE HYDROTREATING

LEGEND

R = REACTOR
F = FRACTIONATOR
HPS = HIGH PRESSURE SEPARATOR
LPS = LOW PRESSURE SEPARATOR
HFS = HOT FLASH SEPARATOR
S = SCRUBBER
CT = CHARGE TANK

UOP 663-12

UOP 663-22

FIGURE 19

FIRST STAGE HYDROTREATING
OPERATIONS PLOT
RUN 1

FIGURE 20

FIRST-STAGE HYDROTREATING
OPERATIONS PLOT
RUN 1

UOP 663-23

FIGURE 21

FIRST-STAGE HYDROTREATING
OPERATIONS PLOT
RUN 2

UOP 663-24

CHARGE STOCK

MAKE-UP H₂

RECYCLE GAS

H₂O

R

HP

LP

S

GASEOUS PRODUCT

LIQUID PRODUCT

SPENT WATER

LEGEND
 R — REACTOR
HP — HIGH PRESSURE SEPARATOR
 S — STRIPPER COLUMN
LP — LOW PRESSURE SEPARATOR

UOP 514-8
UOP 525-77
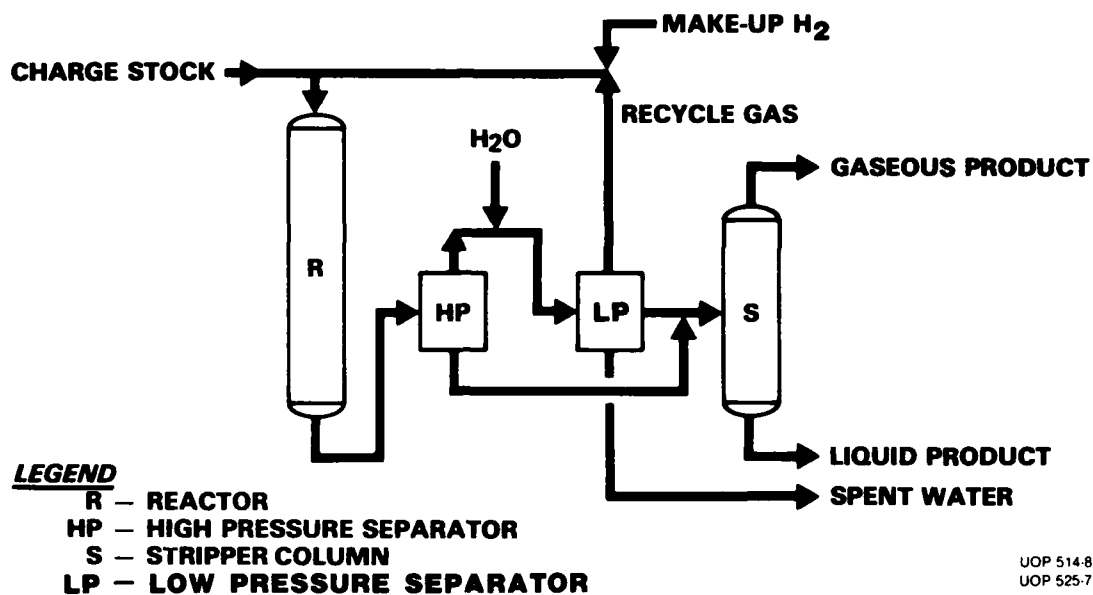
FIGURE 22

HIGH PRESSURE HYDROTREATING
PILOT PLANT
SCHEMATIC FLOW DIAGRAM
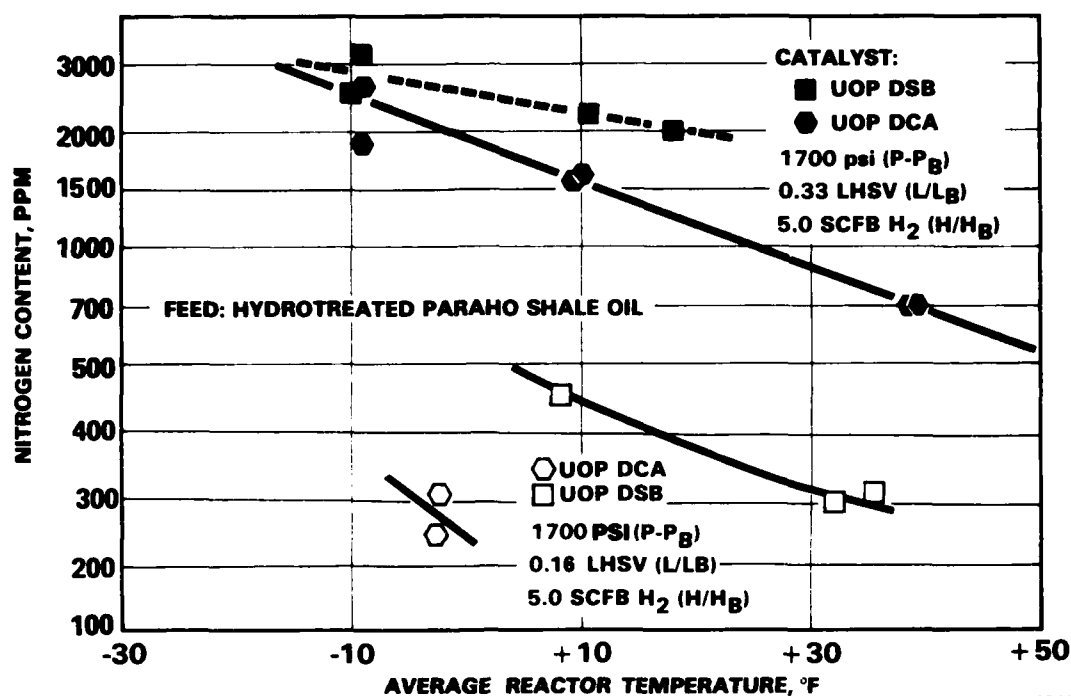
FEED: HYDROTREATED OCCIDENTAL SHALE OIL, 1.1 WT-% N

Legend:
- ■ UOP BCA
- ⬢ UOP DSB
- □ UOP DSA
- ⬡ UOP DCA
- ○ 1ST UOP DCC RUN
- ● 2ND UOP DCC RUN
- △ UOP DCD

1700 PSI $(P-P_B)$
0.33 LHSV (L/L_B)
5.0 SCFB H_2 (H/H_B)
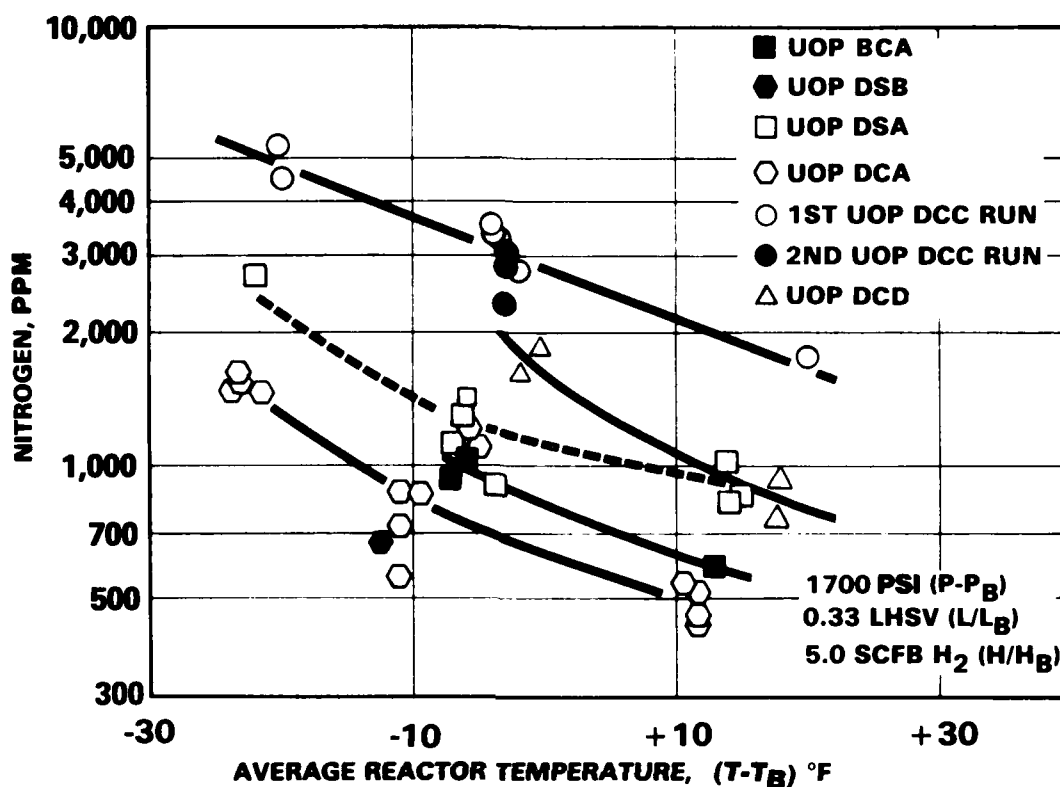
UOP 525-39
UOP 663-27

FIGURE 23

PRODUCT NITROGEN vs.
AVERAGE REACTOR TEMPERATURE

FEED: HYDROTREATED OCCIDENTAL SHALE OIL, 1.1 WT-% N
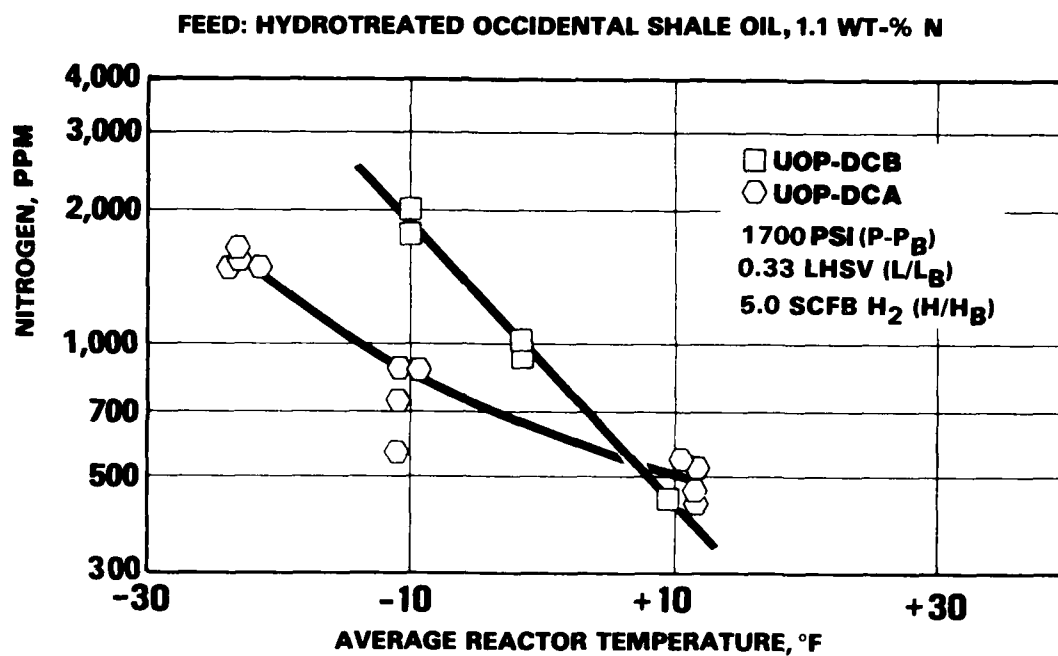
FIGURE 24

PRODUCT NITROGEN vs.
AVERAGE REACTOR TEMPERATURE

FIGURE 25

**PRODUCT NITROGEN vs.
AVERAGE REACTOR TEMPERATURE**
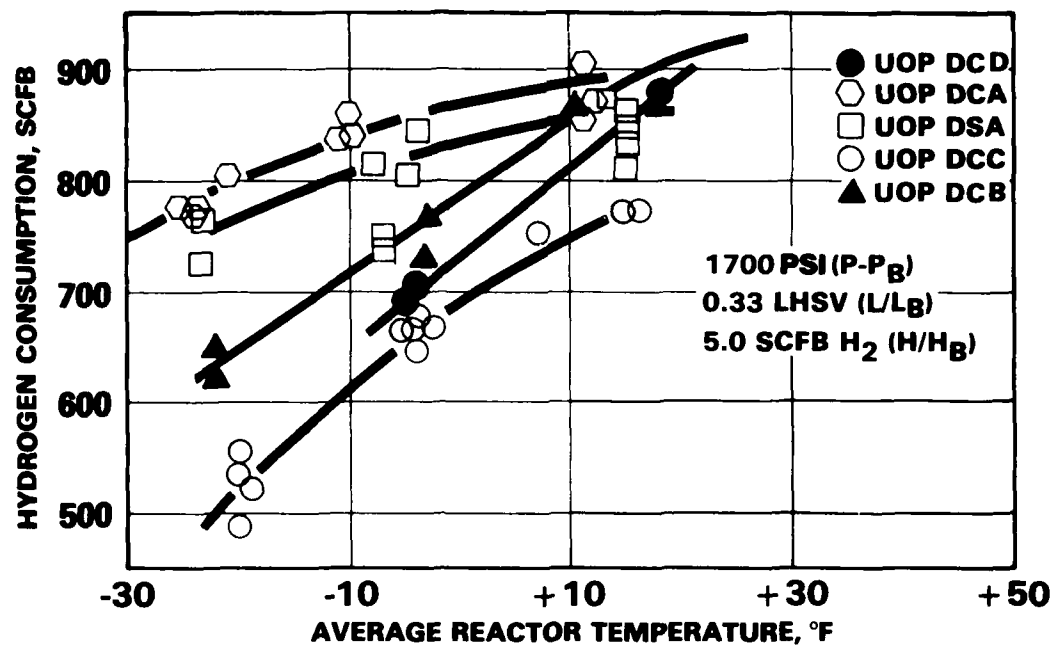
FIGURE 26

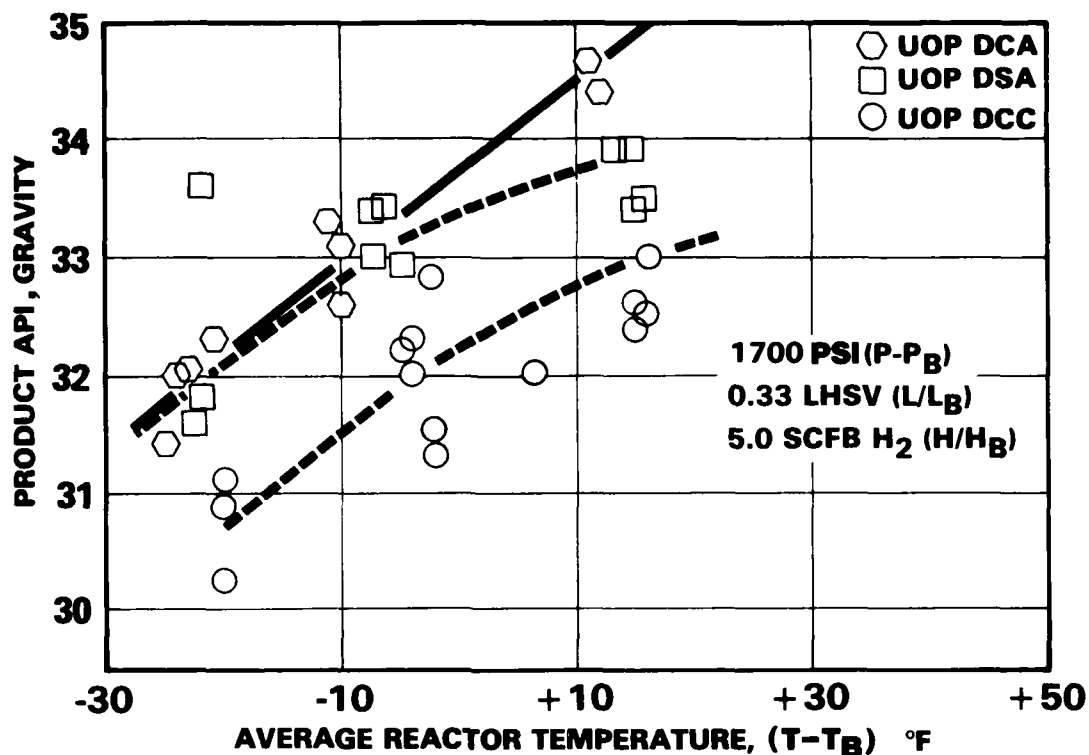## HYDROGEN CONSUMPTION vs. AVERAGE REACTOR TEMPERATURE

FEED: LOW PRESSURE HYDROTREATED OCCIDENTAL SHALE OIL, FEED API 26.6, NITROGEN 1.10 WT-%
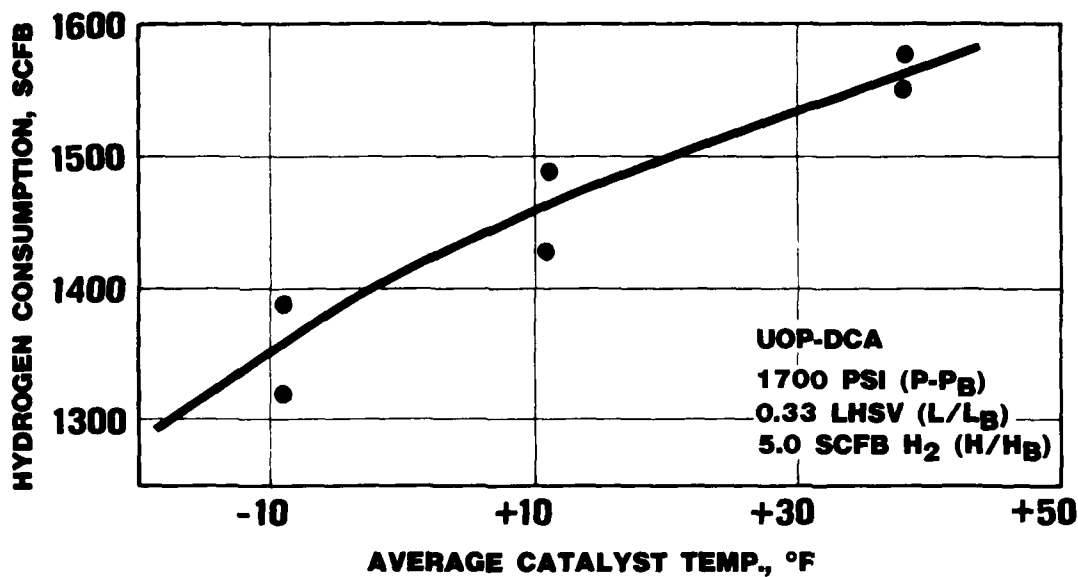
UOP 663-30

FIGURE 27

HYDROGEN CONSUMPTION vs AVERAGE REACTOR TEMPERATURE

FEEDSTOCK: LOW PRESSURE HYDROTREATED PARAHO SHALE OIL

FIGURE 28

PRODUCT API vs.
AVERAGE REACTOR TEMPERATURE

FEED: LOW PRESSURE HYDROTREATED OCCIDENTAL SHALE OIL, FEED API 26.6, NITROGEN 1.0 WT-%
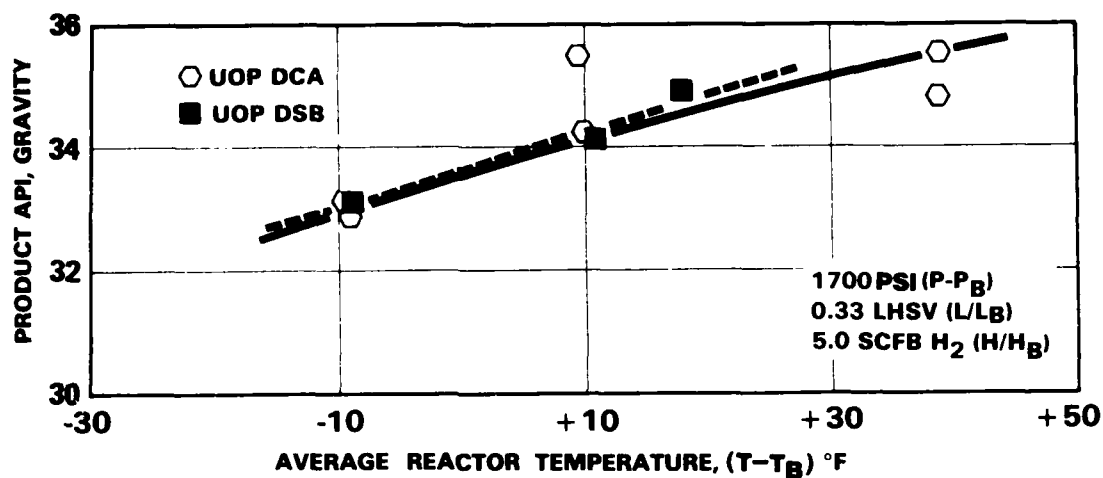
UOP 525 49

FIGURE 29

PRODUCT API vs AVERAGE
REACTOR TEMPERATURE
FEED: HYDROTREATED PARAHO SHALE OIL
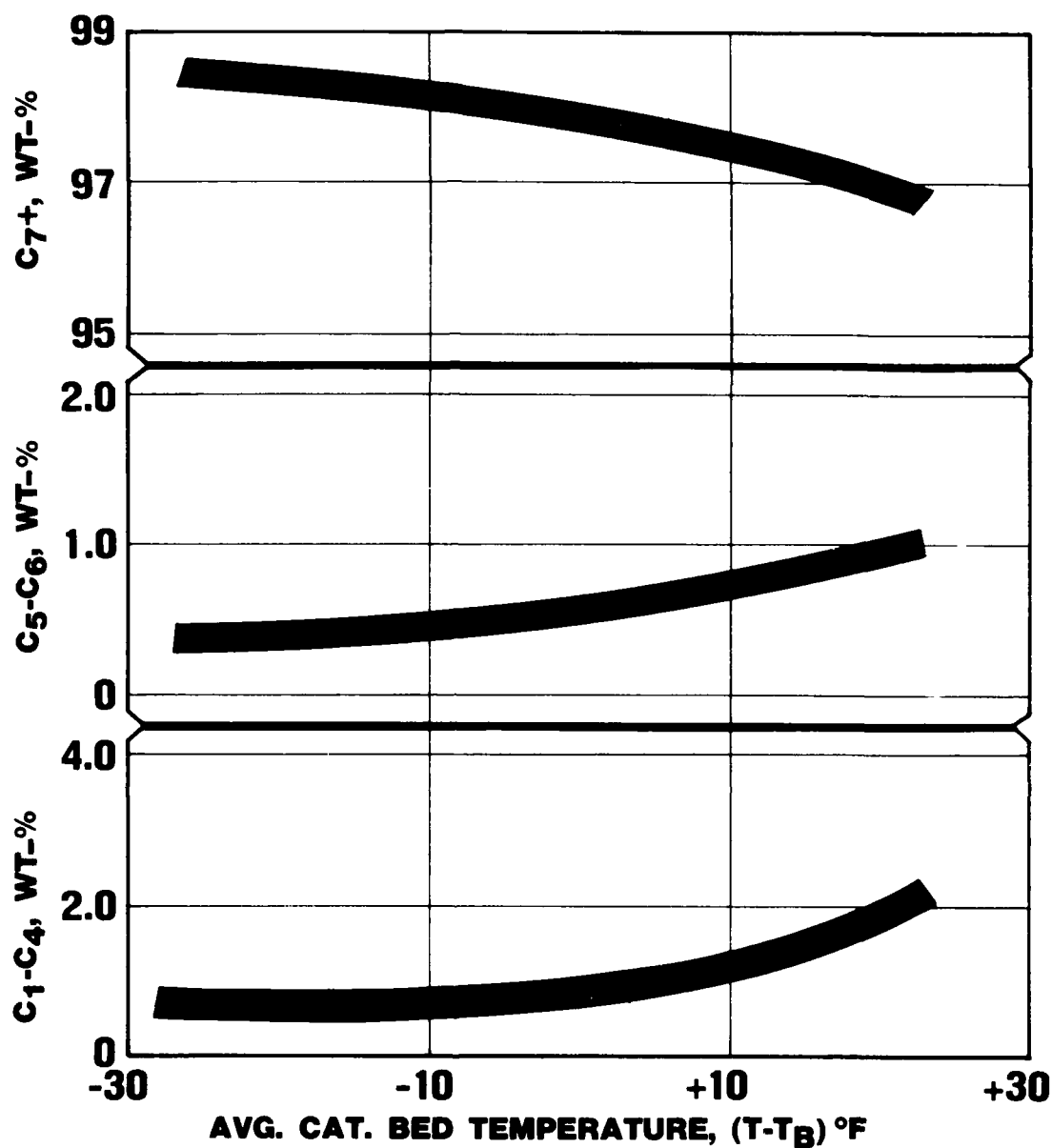
UOP 663-43

FIGURE 30

SECOND-STAGE HYDROTREATING
YIELDS vs. AVG. CAT. BED TEMPERATURE

FEED: HYDROTREATED OCCIDENTAL SHALE OIL
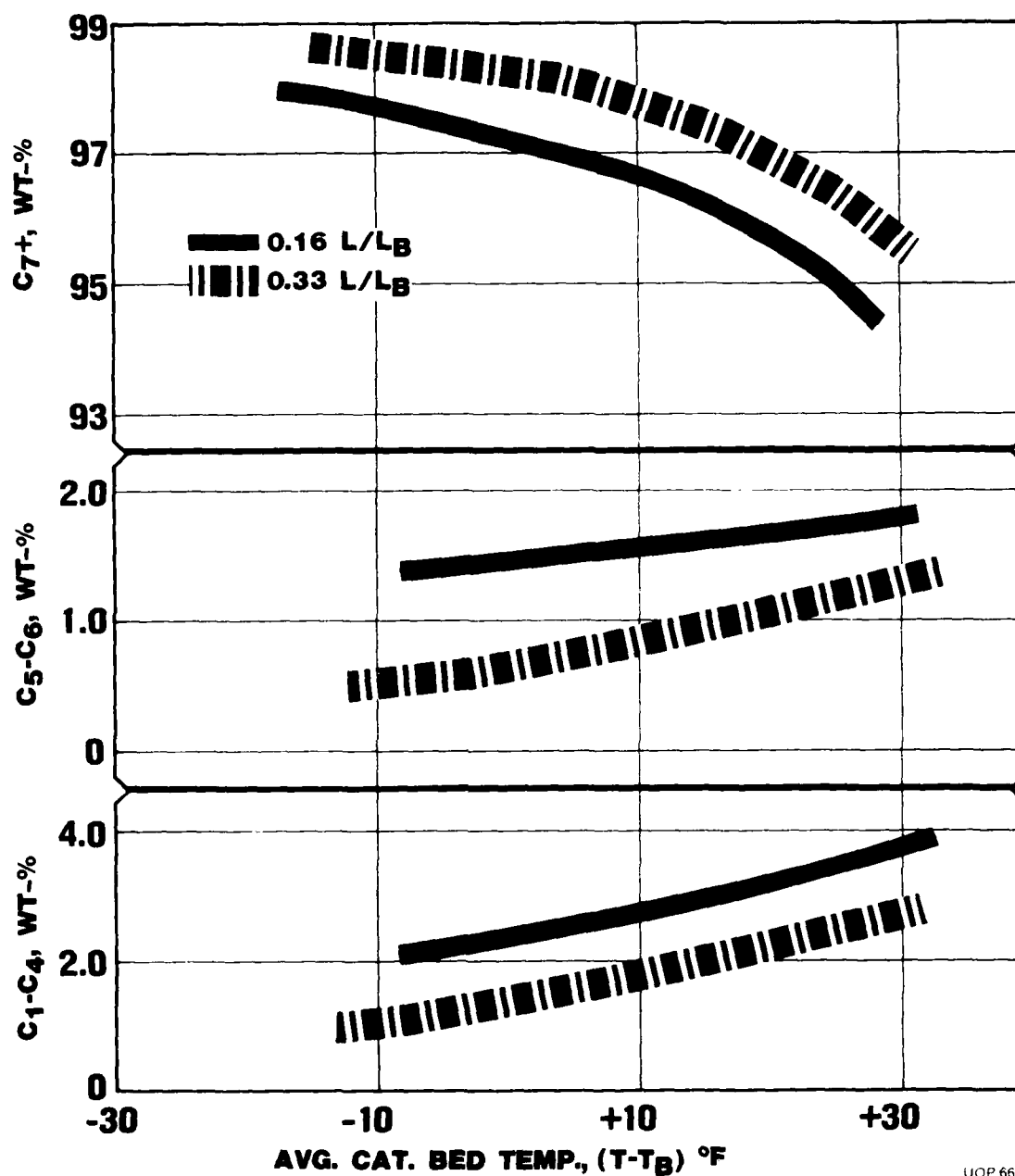
FIGURE 31

SECOND-STAGE HYDROTREATING
YIELDS vs. AVG. CAT. BED TEMPERATURE

FEED: HYDROTREATED PARAHO SHALE OIL

UOP 663-44

CATALYST: UOP-DCA
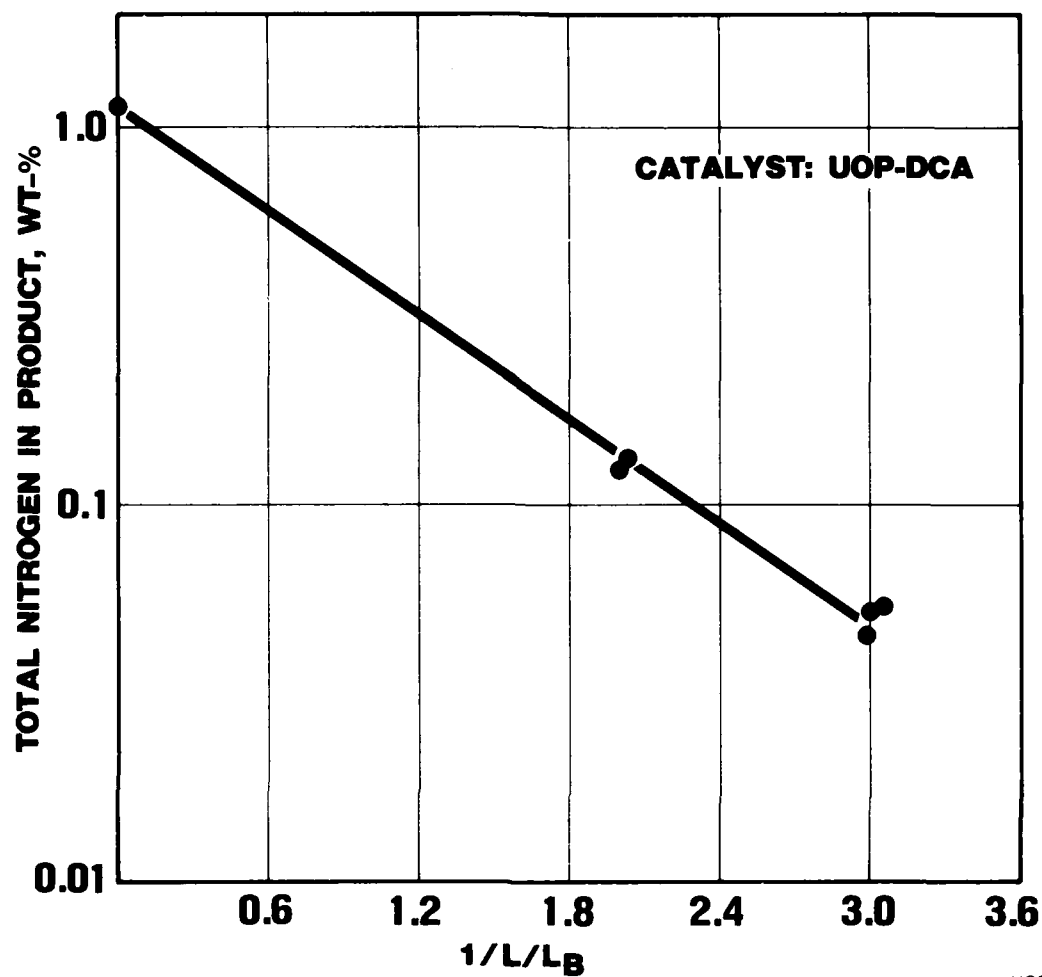
UOP 663-41

FIGURE 32

SECOND STAGE HYDROTREATING
TOTAL N IN PRODUCT vs 1/LHSV
FEED: LOW PRESSURE HYDROTREATED
OCCIDENTAL SHALE OIL
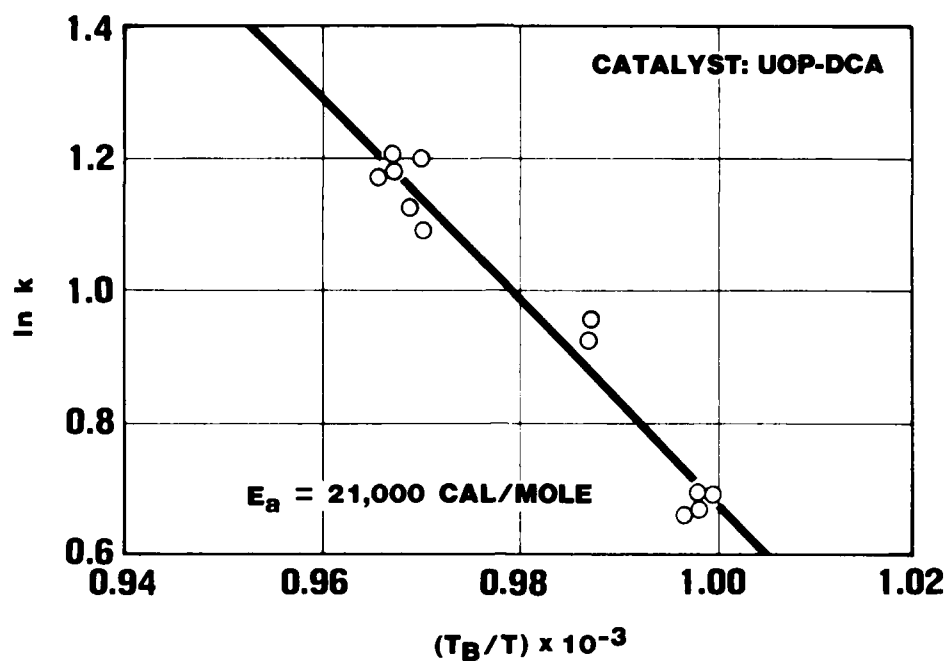
FIGURE 33

SECOND STAGE HYDROTREATING
ARRHENIUS PLOT FOR NITROGEN REMOVAL
FEEDSTOCK: LOW PRESSURE HYDROTREATED
OCCIDENTAL SHALE OIL
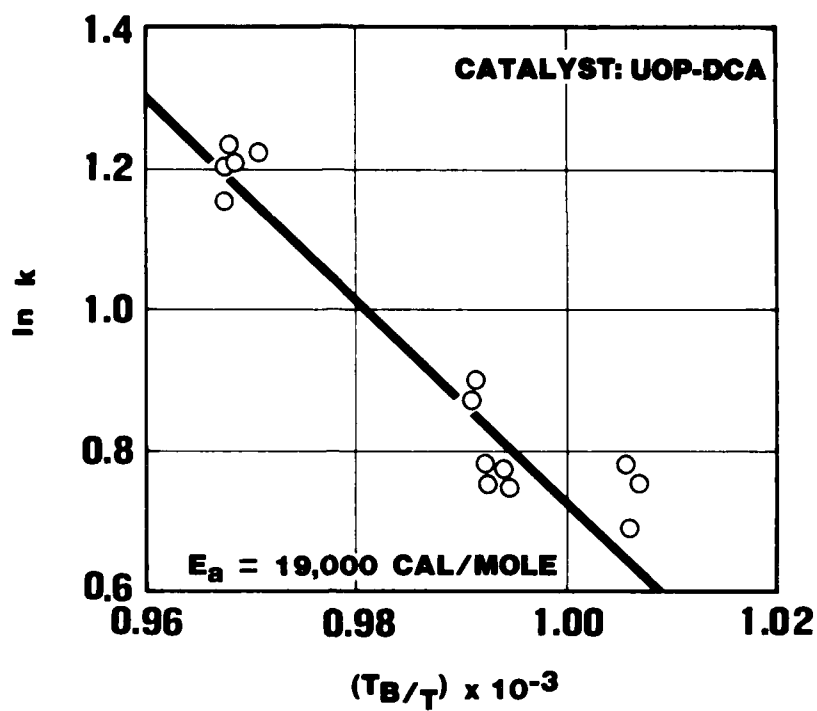
FIGURE 34

SECOND STAGE HYDROTREATING
ARRHENIUS PLOT FOR NITROGEN REMOVAL
FEEDSTOCK: LOW PRESSURE
HYDROTREATED PARAHO SHALE OIL

-147-
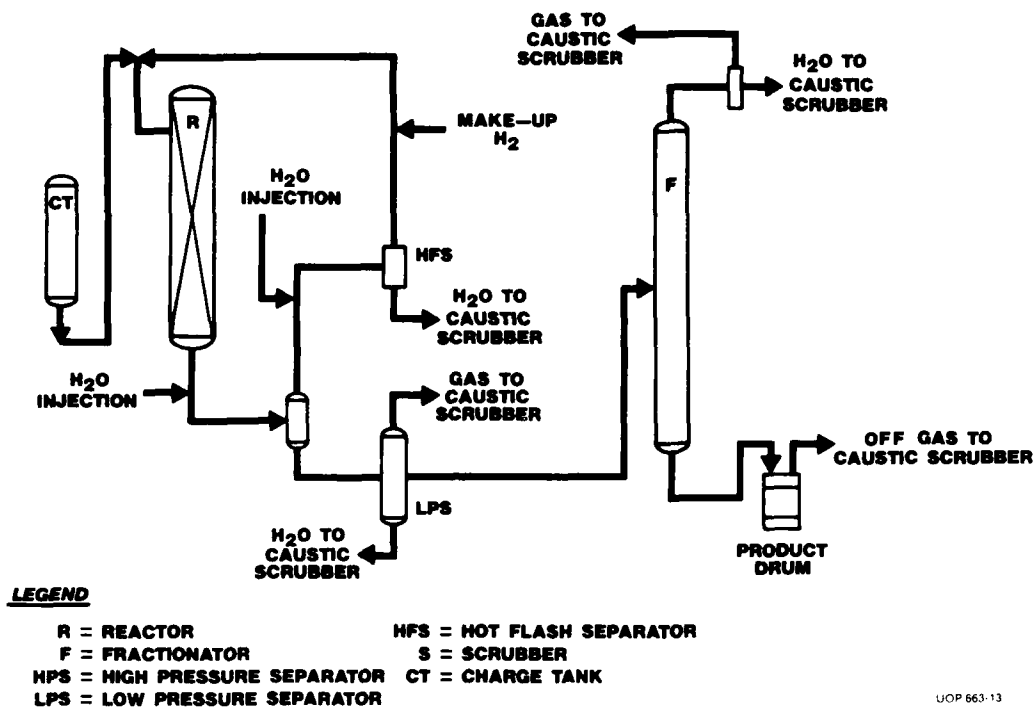
FIGURE 35

PILOT PLANT SCHEMATIC FLOW DIAGRAM
SECOND-STAGE HYDROTREATING
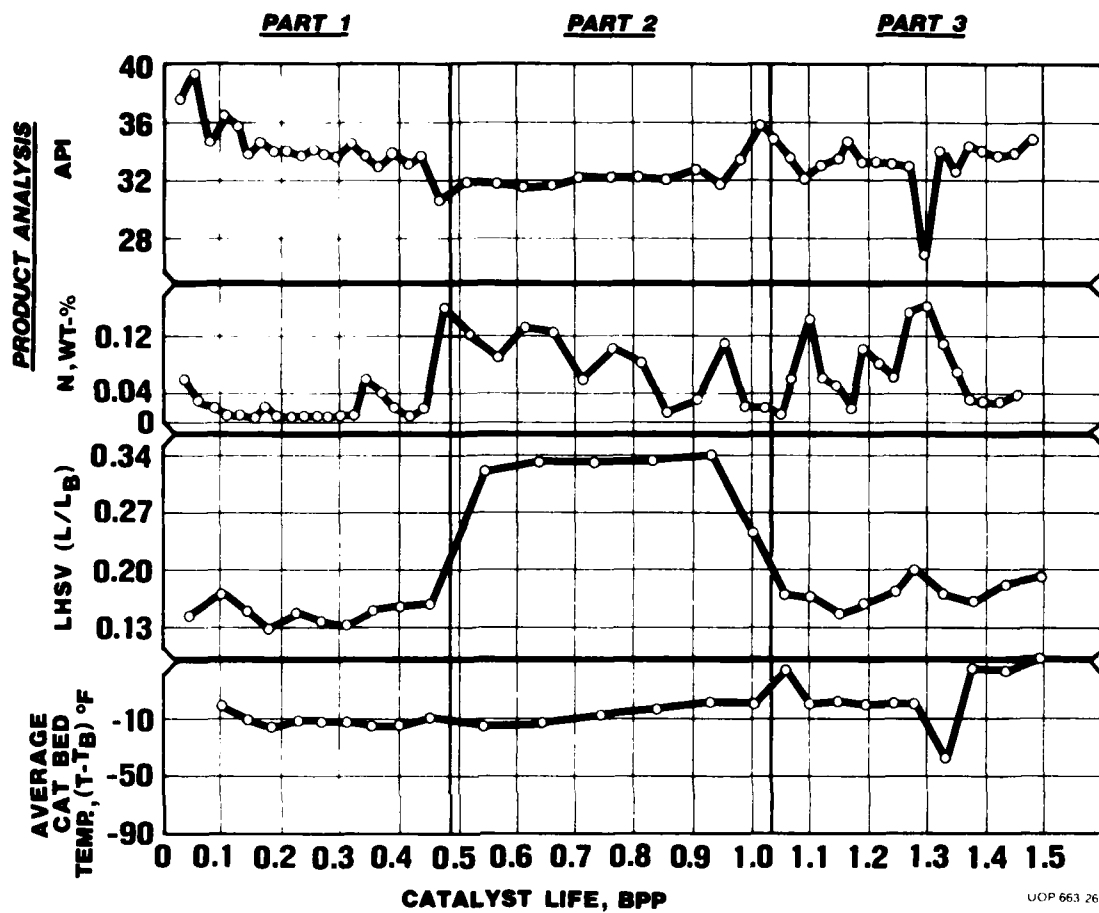
FIGURE 36

SECOND-STAGE HYDROTREATING
OPERATION PLOT
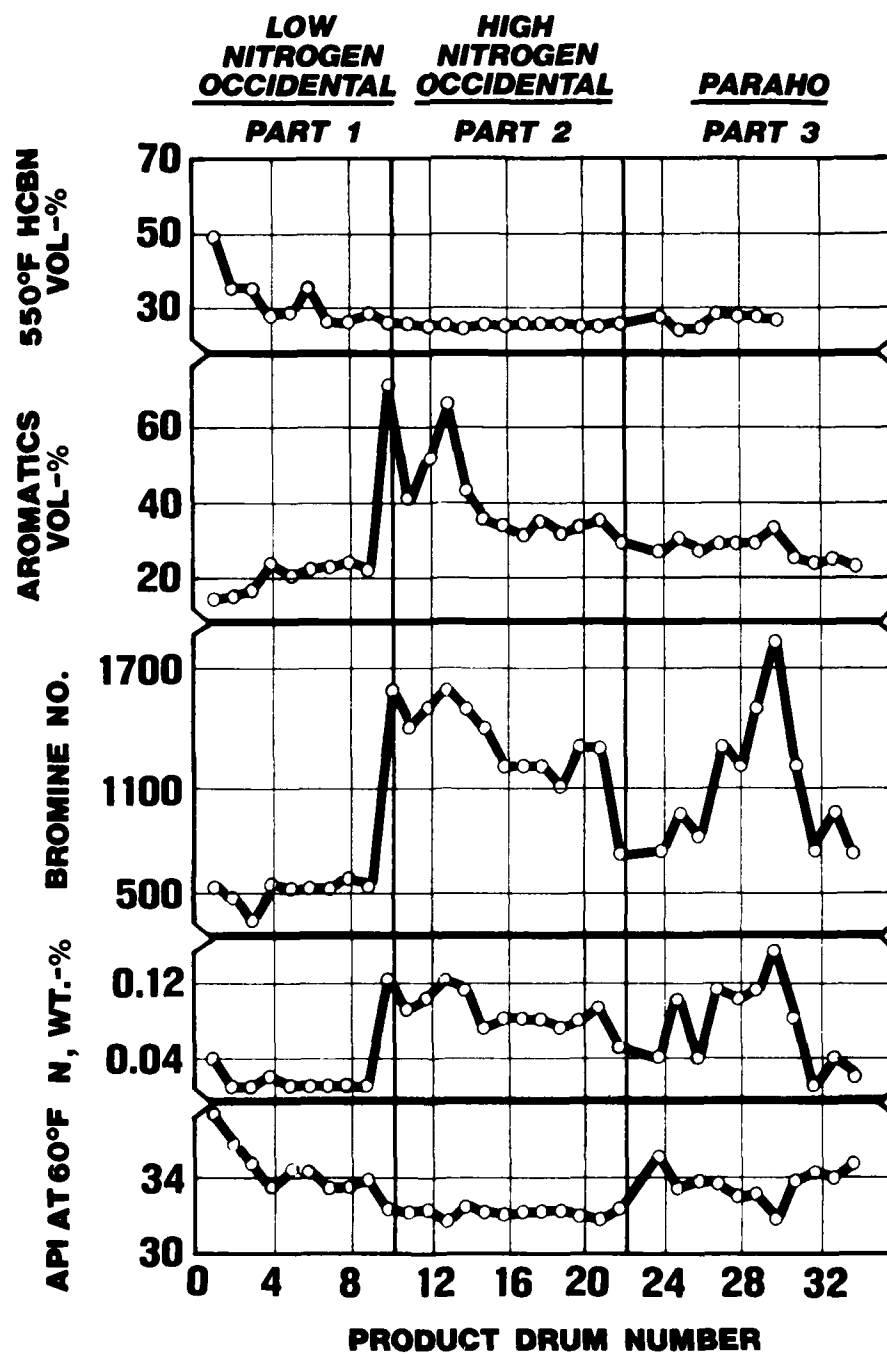
FIGURE 37

SECOND-STAGE HYDROTREATING
PRODUCT DRUM ANALYSIS PLOT

-150-

MAKE-UP H₂

FRESH CHARGE

R = REACTOR
HPS = HIGH PRESSURE SEPARATOR
D = DEBUTANIZER
F = FRACTIONATOR

RECYCLE GAS

R

HPS

D

F

C_4^-

C_5^+ PRODUCT

RECYCLE LIQUID
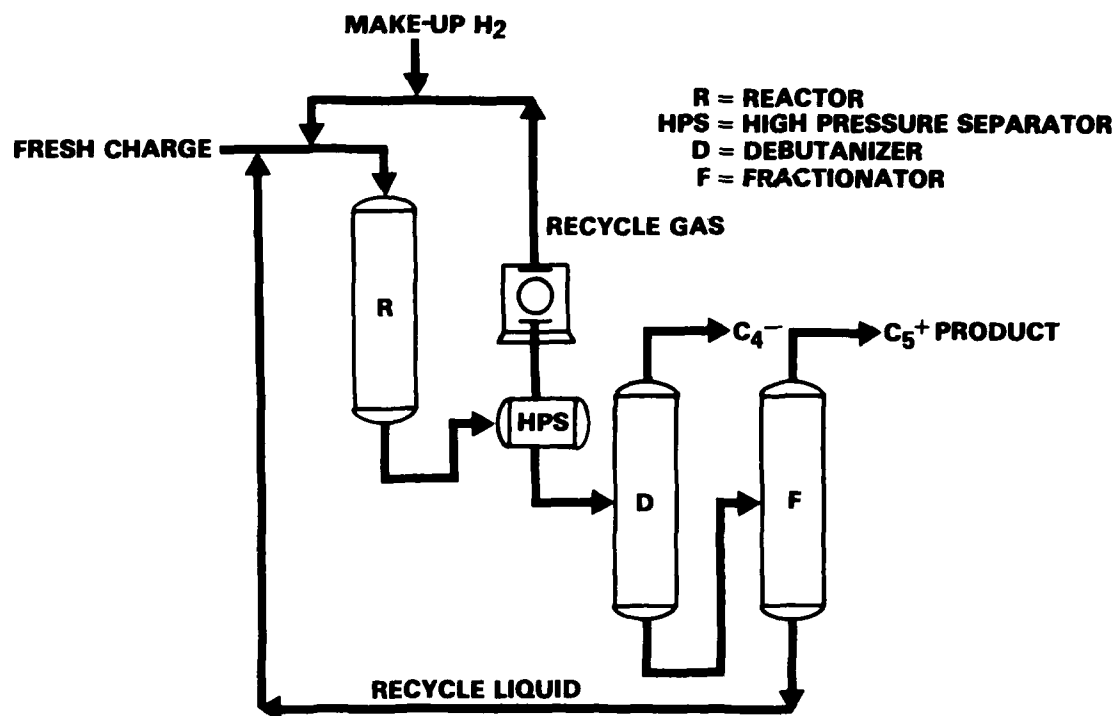
UOP 525-21

FIGURE 38

PILOT PLANT SCHEMATIC –
SINGLE-STAGE HYDROCRACKER

Y-axis: REACTOR AVERAGE BED TEMPERATURE, °F
+50, +30, +10, -10, -30

SOR = —10°F

FEED NITROGEN: 900 ppm
PRODUCT: JP-8
-150 PSI (P-P$_B$)
3.0 LHSV (L/L$_B$)
1.0 CFR (C/C$_B$)
1.20 SCFB H$_2$ (H/H$_B$)

X-axis: 0 50 100 150 200 250 300 350 400 450 500
HOURS ON STREAM
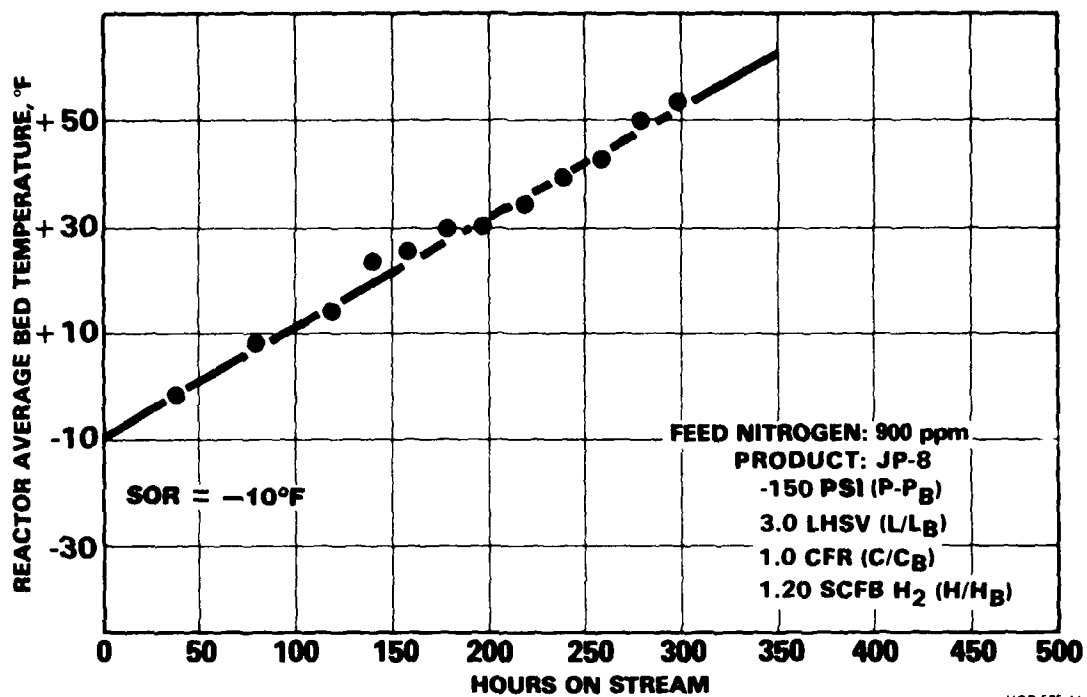
UOP 525-44

FIGURE 39

SINGLE-STAGE HYDROCRACKING
FEED: OCCIDENTAL SHALE OIL

FEED NITROGEN: 900 ppm
PRODUCT: JP-8
-150 PSI(P-P$_B$)
3.0 LHSV (L/L$_B$)
1.0 CFR (C/C$_B$)
1.20 SCFB H$_2$ (H/H$_B$)

SOR = −10°F
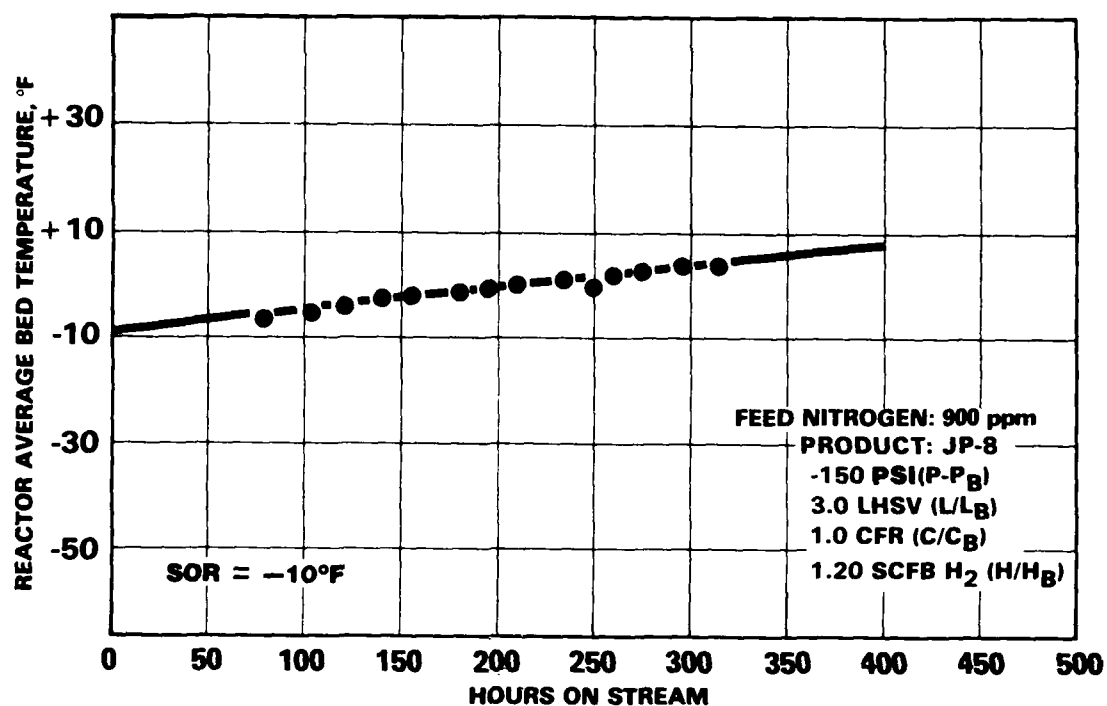
UOP 525 45

FIGURE 40

MODIFIED FLOW HYDROCRACKING
FEED: OCCIDENTAL SHALE OIL

+30

+10

−10

−30

−50

REACTOR AVERAGE BED TEMPERATURE, °F

SOR = −10°F

FEED NITROGEN: 900 ppm
PRODUCT: JP-8
−150 PSI (P-P$_B$)
3.0 LHSV (L/L$_B$)
1.0 CFR (C/C$_B$)
1.20 SCFB H$_2$ (H/H$_B$)

0   50   100   150   200   250   300   350   400   450
HOURS ON-STREAM

UOP 525-85
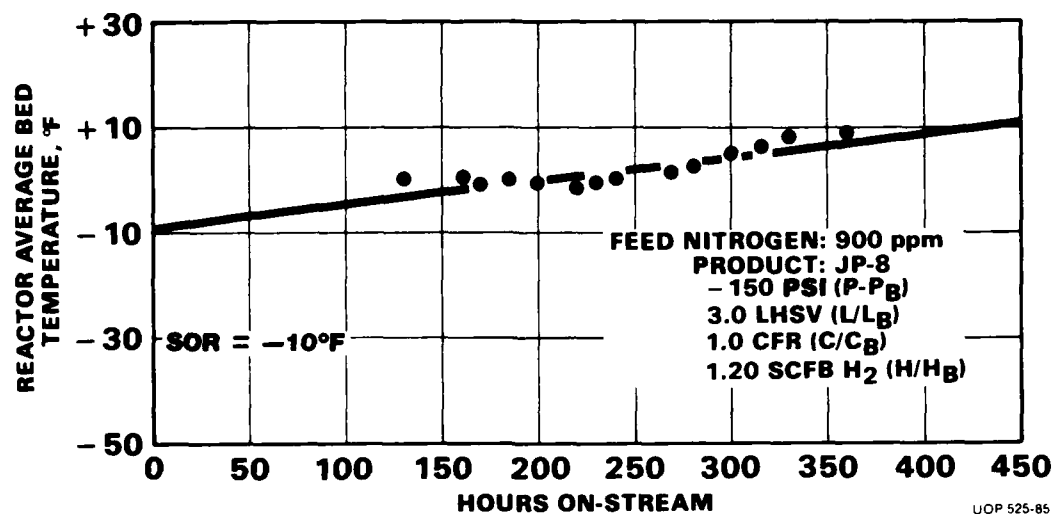
FIGURE 41

PARALLEL FLOW HYDROCRACKING
FEED: OCCIDENTAL SHALE OIL
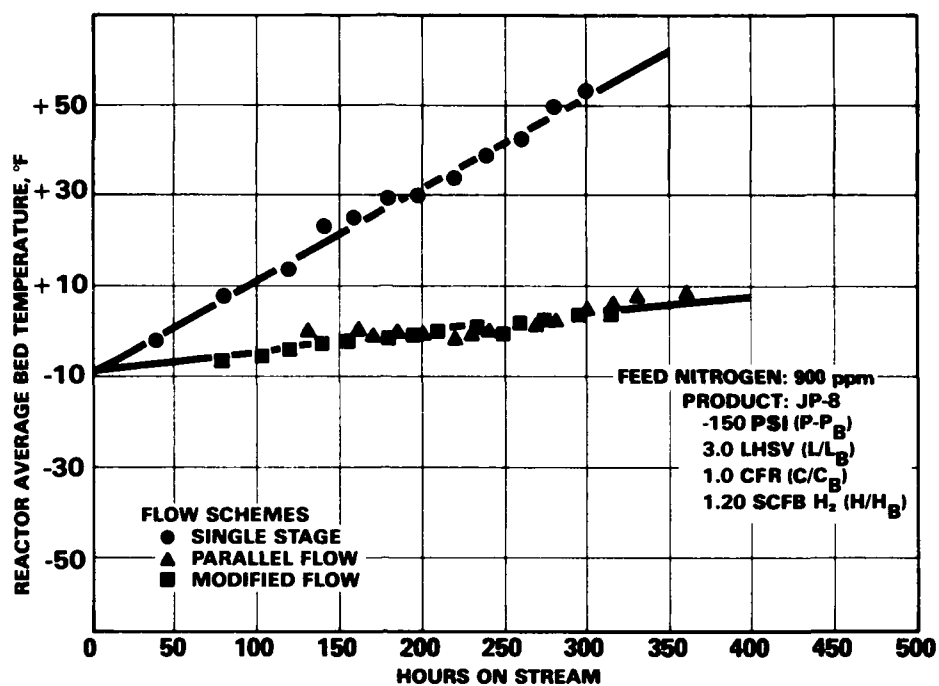
FIGURE 42

COMPARISON OF HYDROCRACKING FLOW SCHEMES
FEED: OCCIDENTAL SHALE OIL

-155-
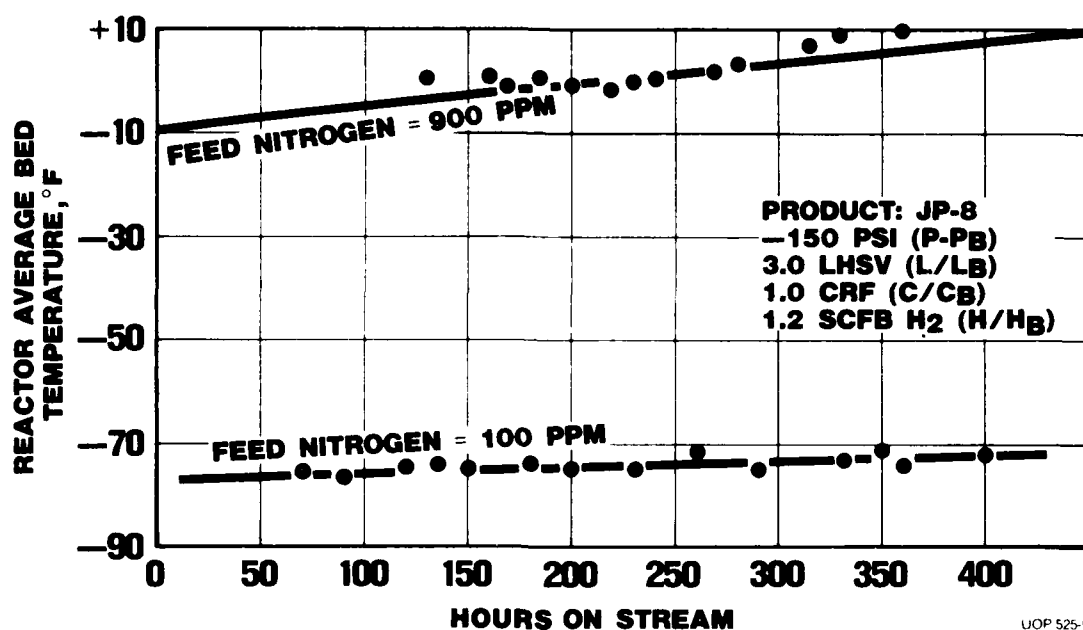
FIGURE 43

PARALLEL FLOW HYDROCRACKING
THE EFFECT OF FEED NITROGEN ON
CATALYST STABILITY

FEED: OCCIDENTAL SHALE OIL

SOR, °F
(T-T_B)

△ PARAHO   -59
○ OCCIDENTAL -71

-150 PSI (P-P_B)
1.4 LHSV (L/L_B)
1.0 CFR (C/C_B)
1.33 SCFB H_2 (H/H_B)
PRODUCT EP = 550°F
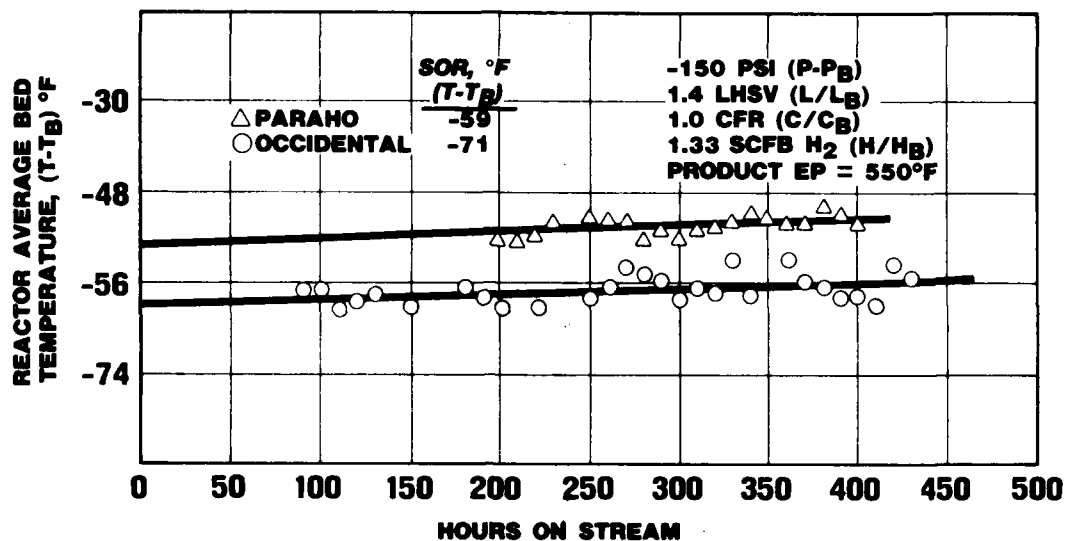
UOP 663-33

FIGURE 44

**PARALLEL FLOW HYDROCRACKING
CATALYST ACTIVITY AND STABILITY**

SECTION III

FOULING STUDY

Introduction

The relative fouling characteristics of shale and petroleum based stocks were evaluated as part of the U.S. Air Force Contract on Shale Oil to Fuels.

The physical and chemical properties of shale oils differ from conventional crude oils, primarily due to the presence of olefins, higher than normal nitrogen and oxygen concentrations, high arsenic, iron and ash contents. The maleic anhydride value for raw shale oil is over 40 signifying the presence of conjugated olefins and styrenes. Since mechanisms of fouling have not been clearly established for petroleum oils, it is apparent that fouling characteristics cannot be predicted, and there is a definite need to experimentally demonstrate the relative fouling characteristics of each specific shale oil.

The objective of this phase of the project, therefore, was the determination of the fouling characteristics of specific shale oils to provide relative fouling rates for comparison to a petroleum derived base material. An existing Monirex® electronic apparatus was used along with specially designed and constructed shale oil circulating equipment to measure fouling rates. The stocks selected for this study included a raw and desalted Occidental shale oil, a raw Paraho shale oil, a raw and desalted Light Berri (Saudi Arabian) petroleum oil, and a kerosine derived from the same crude. The relative fouling characteristics should permit an estimate of a temperature range at which shale oils may be heated before significant fouling would occur.

System Description and Operation

The fouling program involved use of the proprietary UOP Monirex Fouling Monitor (1, 2), using a system modified to handle the viscosity of whole shale oil. The unit directly measures fouling while allowing automatic control of critical parameters such as heat duty, flow rate, pressure, fluid temperature, fouling probe wire (wall) temperature and time at specific conditions. *The technique allows investigation of the effect of* these variables on fouling with various refinery streams. The unit does not accelerate fouling but is electrically sensitive to it.

The system consists of a reservoir, heaters, heat exchangers, filters, pressure regulators, solenoid valve, flow control valve, turbine meter, fouling cell (Figure 45) and associated piping (Figures 46 and 47). The system is contained in a purged, insulated steam-coil- heated box maintaining the fluid at a minimum of 48°C. The fluid would be maintained at room temperature for lower viscosity materials such as kerosine and diesel fuel. The sample handling procedure and equipment for filling and draining the plant have been tailored to eliminate operator exposure to sample. See Appendix A for fill, drain and flush procedures.

The system operates under control of a microcomputer with a permanently programmed memory located in the control cabinet with operator interface via a data terminal (Figure 48). The microcomputer automatically adjusts and maintains the system parameters set for each period. System parameters controlled include fluid temperature to the fouling cell, fluid flow rate, probe wire voltage, the number of periods, period duration and frequency of report. The microcomputer also checks the system for limit conditions, gives an alarm and, if necessary, automatically proceeds with a fail-safe shutdown of the entire system. See Appendix B for microcomputer commands, controls, maximum and minimum parameter limits and alarm messages.

In this configuration (Figure 49), the heated reservoir containing the fluid was sparged with high purity nitrogen to maintain a low oxygen concentration. The major portion of the fluid is recirculated back to the reservoir for mixing, using a gear pump, so that only three gallons of sample are required. A slip stream of 100 mL/minute was regulated through heat exchangers and heaters to the fouling cell. A pressure of approximately 300 psig was maintained at the fouling cell to avoid boiling the test fluid at the elevated probe temperature. From the fouling cell, the fluid was heat exchanged with fluid to the cell and heat exchanged again with cooling water at 39°C before entering the turbine meter and returning back to the reservoir. The turbine meter and associated electronics senses flow rate, compares it to a set point and generates a flow control signal that controls the flow rate via an air operated control valve.

For the purpose of the present tests, the heat exchanger and heaters maintained a constant fluid temperature of approximately 92°C to the fouling cell, while the probe temperature was varied. The heat flux from the heaters to the fluid was low so that no fouling takes place in the heaters. At constant fluid temperature, the probe, initially at approximately 175°C, was increased each period by approximately 25°C increments up to 400°C, and was increased in approximately 50°C increments up to 600°C. The criteria for significant fouling is a rapid decrease in heat transfer coefficient.

The Monirex unit measures current through the iron fouling probe wire at various voltages making it possible to calculate probe temperature. The probe represents a heat exchanger hot metal surface. The change in heat transfer coefficient (h) through the foulant as well as the fouling factor (R_F) can then be calculated. From these data the change in fouling factor may be calculated giving fouling rate (dR_F/dt) at a specific probe temperature.

Fouling is measured directly as a loss of heat transfer:

h = $Q/(T_W - T_{fl})F$, watts/m^2 °C (Btu/hr ft^2 °F)

Q = heat duty, watts (Btu/hr)

F = surface area, m^2 (ft^2)

T_W = wire temperature, °C (°F)

T_{fl} = fluid temperature, °C (°F)

The heat transfer coefficient is converted to a more common unit of the fouling factor (fouling resistance).

R_F = $1/h_t - 1/h_o$, m^2 °C/watt (ft^2 hr °F/Btu)

h_t = heat transfer coefficient at time = t for each period, watts/m^2 °C (Btu/ft^2 hr °F)

h_o = heat transfer coefficient at the beginning of each period, time = 0

The fouling factor is relatively linear over the test period after initial equilibrium of the probe wire temperature, which requires a few minutes. Time periods for these studies were three hours each.

The time differential of the fouling factor can be calculated giving fouling rate (dR_F/dt) and when plotted versus wire temperature using an Arrhenius plot, can be used to determine fouling as a function of heat exchanger wall temperature.

Experimental Data and Discussion

The UOP Monirex Fouling Monitor, therefore, measures the temperature of the fluid (T_{fl}), the wire (T_W) and the heat input (Q) to the wire. Since the area is known, it is possible to calculate the heat transfer coefficient (h) as well as the fouling factor (R_F) and fouling rate (dR_F/dt). Data reported here include all of these values (except wire

area) with maximum and minimum h values for specific temperatures. The wire is a 7.0 cm long, 0.190 mm diameter, high purity iron coiled into a spring approximately 2.5 cm long.

Since oxygen is known to affect the fouling and aging characteristics of complex hydrocarbon mixtures, such as the shale and petroleum oils, all samples received are sparged with nitrogen to remove the dissolved oxygen, blanketed with nitrogen and stored at 40°F (4.4°C) until ready for test. The samples were removed to a 105°F (40.5°C) area for warming for easy transfer to a portable container for transport to the Monirex unit. All samples were continually blanketed with nitrogen.

The samples received for this program were analyzed and these data are given in Tables 79 through 84. The analyses of the petroleum desalted and raw crude oils were provided by the supplier.

The unusual characteristics of the shale oils are the higher density and higher arsenic and nitrogen contents than those normally found in petroleum stocks. The bubble point, diene value, and Bromine Number are also higher for the shale oils. The importance of these characteristics relative to fouling has not as yet been determined and it is not feasible to speculate as to the magnitude of relative fouling rates of these materials. It should be noted that no studies were planned as a part of this project to make these correlations.

The desalted Occidental shale oil was prepared with a commercially available deashing/dewatering unit. The preflash drum removed the last traces of water from the feed and eliminated fouling associated with vaporization of this material in downstream processes. In addition, the treated shale oils were filtered through a 15 micron filter and stored in nitrogen blanketed tanks.

The 100% Light Berri crude was obtained from Texaco. The desalted crude contains a demulsifier, Treatolite DS 636. Both the raw and desalted crude oils contained less than 0.3 ppm oxygen and the refinery indicated that the desalted crude was low in fouling tendency, causing a pressure drop of about 30 psig across the crude unit preheat exchangers after a one year run. Low fouling rates are to be expected with these petroleum stocks.

Fouling test conditions were established in accordance with the expected range of shale oil refining process conditions, or as otherwise limited by capabilities of the Monirex Fouling unit. The flow rate was expected to vary between 300 and 800 mL/min past the wire test probe and the pressure was high enough to prevent boiling at the heated wire. The proposed raw shale and petroleum oils conditions were 200°F (93.3°C) fluid temperature and 325°F (162.8°C) wall temperature. Test conditions for the desalted samples were based on those listed in Table 85. Refinery fouling regularly occurs on the shell side of the heater and heat exchangers leading to a hydrotreater. Therefore, temperatures of 200 to 280°F (93.3 - 137.7°C) for the fluid and 360 to 500°F (182 - 260°C) for the wire were considered. To include all of these test conditions, the original proposed fluid temperatures were 200 to 450°F (93.3 to 232°C) and the wire temperature was 160°F (71.1°C) higher than the fluid temperature.

Most fouling studies reported in the literature involve the passage of a small quantity of liquid over a heated tube surface. In the JFTOT used by Hazlett (3), or the Advanced Kinetic Unit used by Taylor (4), the liquid is passed over a heated stainless steel tube with heated zones that could vary several hundred degrees within a single zone.

The Monirex Fouling unit offers a unique opportunity to make kinetic studies with a large volume of test fluid (offering pseudo first-order kinetics), and a heated iron wire with a small surface area and with a known temperature. With the Monirex unit it is possible to have a controlled fluid temperature and the computerized system permits control and measurement of the wire temperature for predetermined periods of time. It

is possible, therefore, to have a series of temperatures and determine the fouling rate at each temperature. These data can be plotted as the log of the fouling rate vs. 1/K (where K is the absolute temperature), to obtain an Arrhenius plot. Activation energies can be calculated and any significant changes in activation energy are indications of significant changes in the fouling reactions.

The fluid temperature finally selected for this program was 90 - 92°C and the wire temperatures were varied from approximately 175°C to temperatures of 600°C. The low fluid temperature was selected to minimize fouling within the plant, and examination of various components indicated no apparent fouling in the stainless steel components. The pure iron wire probe was heated in 25°C increments from 175 to 400°C and then in approximately 50 - 75°C increments up to 600°C. Each test period was 3 hours long and the fouling rate was calculated for the period 0.5 to 3.0 hours. The test fluid was continually sparged with high purity nitrogen (99.9999% N_2) and the fluid flow was 100 mL/min past the wire probe. The initial, final and mean wire temperature, the fluid temperature, the original and final h values and the dR_F/dt for each period are reported. A regression analysis of the dR_F/dt values for the 0.5 to 3.0 hour periods was made and these fouling rates are given in the tables.

The initial experiments with the Monirex Fouling unit were done with the petroleum kerosine, Runs 1 and 2, Table 86 and Figure 50. Run 1 was a successful experiment. Since the pump failed prematurely in Run 2, the planned temperature cycle was not completed. The kerosine was evaluated at a liquid flow rate of 400 mL/min.

The instrumentation used to measure changes in resistance is a limiting factor in the precision obtained in the fouling rate (dR_F/dt), and it has been found that the values of 1 or less must be evaluated as < 1. The data obtained in Run 1, Periods 4, 5, 6 and 7 and Periods 6, 7 and 8 in Run 2 must be estimated as fouling rates of < 1 and cannot be used in making an Arrhenius plot. These data suggest that the Texaco

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kerosine fouls at low rates (< 1), at temperatures less than the 225 - 235°C range. The data useful in making the Arrhenius plot, therefore, fall in the 235 - 400°C temperature range and these data are shown (*) in Table 86 and in Figure 50.

The Light Berri kerosine appears to be a low fouling product, a conclusion that has been confirmed by a verbal report from Texaco. The apparent activation energy is -8.5 kcal and is in the range for JP fuels reported by Taylor.

The kerosine had been previously evaluated with the original Monirex unit and the dR_F/dt values for a fluid temperature of 150°C and a wire probe temperature of 300°C ranged from 1.3 to 2.0 for the 1 to 4 hour exposure period. The average value from Figure 50 for the same kerosine is 3.7. This is a reasonable correlation between the two sets of data and the difference is most probably due to the difference in fluid temperature.

The desalted Light Berri crude petroleum oil was also evaluated at a flow rate of 400 ml/min. and these data are summarized in Table 87 and Figure 51. Only the data obtained in Periods 4B through 7B have been used to make the Arrhenius plot. Run 4 was aborted due to instrument failure. This oil does not appear to foul at a significant rate at temperatures below 200 - 225°C and has an apparent activation energy of -8.5 kcal.

Since the flow rate of 400 mL/min was significantly higher than that used in the original fouling apparatus, the flow rate was reduced to 100 mL/min in all subsequent experiments. The desalted crude oil was again evaluated and these data are shown in Table 88 and Figure 52 (Run 5). Reducing the flow rate resulted in a significant increase in the precision of the experiment and now six periods have been used to make the Arrhenius plot. These data confirm that obtained in Run 3 in that there is no significant fouling at temperatures below 200°C. At temperature greater than 400 - 450°F, the fouling rate as measured by this technique shows a

significant difference from that found in the temperature range of 200 - 400°C. This may be due to a change in the chemistry of thermal degradation of the oil on the wire, or to mechanical aspects such as a very thick film on the wire or perhaps some of the foulant actually broke off the wire. Again data below 200 and above 400°C were not used in the Arrhenius plot. These data show that the apparent activation energy is similar to that for the kerosine and that the crude oil does foul at a greater rate and at a lower temperature than the kerosine.

The relative antifouling characteristics of the raw crude oils with and without a proprietary antifoulant are shown in Tables 89 and 90 and in Figure 53. The raw crude oil has a slightly lower fouling rate than the desalted crude and the addition of the antifoulant had no effect on the fouling rate.

The desalted Occidental shale oil was evaluated in duplicate, and these data are summarized in Table 91 and Figure 54 (Runs 7 and 9). The temperature at which significant fouling occurs is 175 - 200°C and this shale oil fouls at a rate greater than the crude. The apparent activation energy is -4.6 kcal which is significantly lower than that obtained with the Light Berri crude oil. At temperatures of 400°C or higher, there is a significant difference in the fouling rates than those predicted from the 200 to 400°C range, again suggesting either an experimental artifact or that there is indeed a different mechanism by which fouling occurs at the higher temperatures.

The relative antifouling properties of the raw Occidental shale oil are shown in Table 92 and Figure 55. The data obtained with the desalted Occidental shale oil is also shown in Figure 55. These data show that the raw shale oil fouls at a higher rate than the desalted shale oil. The slopes of the two curves are also similar indicating similar activation energies. The shale oil does foul at a higher rate than the petroleum crude, although it appears that even this shale oil sample fouls at a relatively low rate.

The desalted Paraho shale oil, with and without an antifoulant, was evaluated and the relative antifouling properties are shown in Tables 93 and 94 and Figure 56. The Paraho oil does not follow a normal single Arrhenius plot. There appear to be two possible reactions. The first dominant reaction occurs in the temperature range of 170 to 315°C (1/K of 0.00226 to 0.0017), with an apparent activation energy of +2.8 kcal and the fouling rate actually decreases with increasing temperature. Taylor has noted that similar decreases in fouling rate are due to the presence of polysulfides or oxygenates (4).

Although the Paraho oil was stored in 5-gallon cans and purged with nitrogen, the seal on these cans is poor and it is not unreasonable to assume that oxygen diffused into the shale oil. The Paraho oil had been pre-aged under these conditions for approximately one year before evaluation for fouling characteristics and, under these conditions, it is possible to form disulfides and oxygenates.

At temperatures higher than 315°C, the Arrhenius plot (apparent activation energy of -11.3 kcal) is higher than that obtained with the desalted Light Berri crude petroleum oil or the Occidental shale oil and the fouling rate increases with temperature. It is also possible that the Paraho shale oil fouls at such very high rates so as to deposit sufficient foulant on the iron wire to change the wire dimensions and results in erroneous h calculations. Sufficient Paraho shale oil was not available to check this possibility.

The addition of a proprietary antifoulant at 100 ppm to the Paraho shale oil had a significant effect on the fouling rate at temperatures to 280°C and no effect at higher temperatures. Only the initial reaction was influenced by the antifoulant. Although no comments can be made as to the chemistry of this antifouling action, it is clearly evident that the action of an antifoulant can be measured with the Monirex Fouling unit and that the fouling rate was reduced at temperatures up to 280°C.

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The activation energies (AE) of various products are shown in Table 95. The raw Light Berri crude oil with and without antifoulant have lower AE values than the desalted crude oil and this is the same trend observed with the raw and desalted Occidental shale oils. No data are available to explain these differences, although it can be speculated that desalting removes components that affect the fouling reactions. It is also apparent that the addition of an antifoulant resulted in a small reduction of the AE for both the raw crude oil and the desalted Paraho shale oil. Finally, at temperatures higher than 315°C, the desalted Paraho shale oil had the highest AE.

The linear regression calculations based on the Arrhenius plots were used to calculate fouling rates in the 200 to 400°C temperature range. Since the correlation coefficients for the linear regression calculations range from 0.95 to 0.99, these calculations are a good indication of average experimental values. The relative fouling rates are summarized in Tables 96, 97 and 98 and in Figures 57 and 58. These data show:

1. Decreasing the flow of the desalted crude oil from 400 to 100 mL past the iron wire probe had no significant effect on the fouling rate data.

2. The desalted crude oil fouled at a higher rate than kerosine derived from this oil. The original crude oil, however, had an unusually low fouling rate and it is suspected that these low values are due to the presence of the processing additives. This theory is supported by the lack of activity of a proprietary antifoulant added to the crude oil, and the lower activation energy of the crude when compared to the desalted crude and the kerosine.

3. The raw Occidental shale oil fouls at a greater rate than the desalted shale oil. At the lower temperatures (200 to 360°C), the desalted petroleum crude fouls at a lower rate than the Occidental desalted shale oil, although at temperatures of 360°C and higher, the oil fouls at a higher rate than the Occidental shale oil.

4. At the lower temperatures, from 200 to 250°C, the Paraho desalted shale oil appears to foul at a higher rate and, at temperatures greater than 250°C, fouls at a lower rate than the Occidental desalted shale oil. Although it appears that the Paraho desalted shale oil should not present any significant fouling problems in this temperature range, the unusual shape of the Arrhenius plot makes the data suspect. These data should be rechecked with a fresh Paraho shale oil sample with a known aging history. The addition of the proprietary antifoulant reduced the fouling rate of the Paraho deashed shale oil in the 200 to 300°C temperature range but had no effect in the 300 to 400°C range.

5. The activation energy values indicate different fouling characteristics for the Light Berri petroleum fractions and the Occidental and Paraho oils. This is consistent with the differences reported in the concentration of those compounds suspected of fouling reactions.

In the original test procedure utilizing increasing wire temperatures, two major assumptions are made. It was assumed that the wire surface area does not change significantly and that the initial wire temperature is the surface temperature. To evaluate these assumptions a series of experiments was made with the probe voltage relatively constant for an extended period of time. The maximum temperature of general interest is approximately 350°C (700°F), and experiments were made with the Occidental desalted shale oil with initial probe temperatures of 250, 280 and 310°C. A new wire probe was used for each experiment and the probe voltage remained constant for periods up to 42 hours. These data are summarized in Table 99 and 100 and Figure 59. The Arrhenius plot obtained with the Occidental desalted shale oil using the original step procedure is also shown in Figure 59.

The data obtained with a constant applied voltage (constant surface temperature) is similar to that obtained with the original procedure employing increasing wire temperatures. A linear regression of these data yielded curves similar to that obtained originally and it is apparent that the original assumptions are correct. High fouling rates, however, may lead to errors, and long test periods at higher fouling rates should also be avoided.

As a final assessment of these experimental Monirex Fouling unit procedures, two experiments were made with the Occidental desalted shale oil to show the effect of increasing and decreasing the wire probe temperature on dRF/dt. These data are summarized in Table 101 and Figure 60. Although sufficient data are not available to make accurate Arrhenius plots, the data show that different fouling rates were obtained with successive increasing and then decreasing wire probe temperatures. It appears that the increasing wire probe sequence should be limited to low fouling materials and that high fouling materials should be evaluated with a new wire probe for each temperature.

Conclusions

The Monirex Fouling unit has been successfully used to evaluate the relative fouling rates of a raw and desalted Light Berri petroleum oil, a raw and desalted Occidental shale oil, and a desalted Paraho shale oil, as well as the effect of a proprietary antifoulant on the fouling rate of the raw crude and the desalted Paraho shale oil. These data were obtained by increasing an iron wire probe temperature in approximately 25-50°C increments from 175 to 400°F and then in 100°C increments to 600°C. It was possible to make Arrhenius plots and to calculate relative fouling rates. Two other test procedures -- the application of a constant voltage for an extended period of time, and a rapid increase and decrease in wire temperature -- confirmed the usefulness of the original test sequence.

The desalted crude petroleum oil fouled at a higher rate than the kerosine derived from this oil. The original raw crude, however, had a lower fouling rate than the desalted crude and it is suspected that this

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is due to the presence of processing additives. The raw Occidental shale oil fouled at a greater rate than the desalted shale oil. At lower temperatures (200 to 300°C), the desalted petroleum crude fouls at a lower rate than the desalted Occidental shale oil; although at temperatures of 360°C and higher, the petroleum oil fouls at a higher rate than the Occidental shale oil. However, all of these samples fouled at relatively low rates.

Paraho shale oil appears to have different fouling properties than the Occidental shale oils with higher fouling rates at temperatures up to 250°C and with lower fouling rates at temperatures higher than 250°C. These data are suspect, however, and a fresh Paraho shale oil should be evaluated to eliminate the probability that high fouling rates confounded the present experimental technique. A proprietary inhibitor does reduce the fouling rate of the Paraho shale oil at temperatures less than 280°C.

Although the data from this initial investigation of the relative fouling tendencies suggest that these particular shale oil samples foul at relatively low rates, additional information is required to confirm this. In particular, the effect of sample handling (aging) and the detailed analysis of pilot plant performance must be completed. Both of these concerns are to be addressed in subsequent phases of this contract.

FOULING STUDY REFERENCES

1. R. Braun and R. H. Hausler, Oil & Gas J., 75, 6, 90 (February 14, 1977).

2. R. Braun, Materials Performance, 16, 11, 35 (1977).

3. R. H. Hazlett, et al., Ind. Eng. Chem. Prod. Res. Dev., 16, 2, 171 (1977).

4. W. F. Taylor, et al., Ind. Eng. Chem. Prod. Res. Dev., 17, 1, 86 (1978).
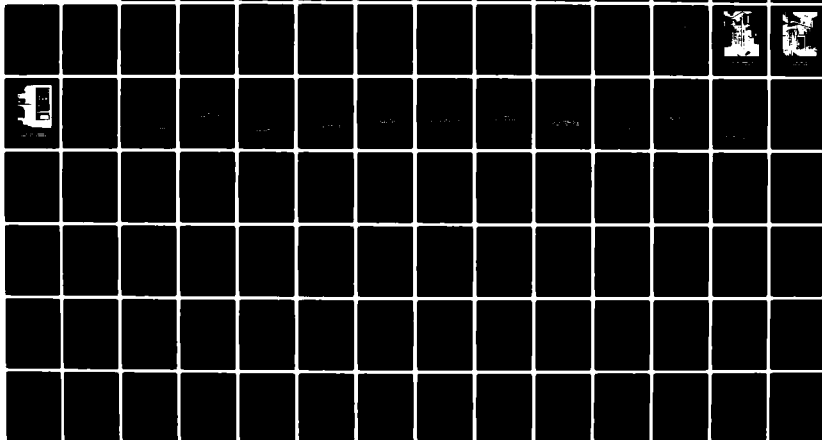
TABLE 79. DESALTED OCCIDENTAL SHALE OIL

UOP No. 49-1098A

		Metal by Emission, ppm	
API Gr. at 60°F	22.9		
Sp. Gr. at 60°F	0.9165		
Bubble Pt. at 500 psig	1056°F	Fe	42
Distillation, °F		Mn	< 0.2
IBP	376	Cr	0.21
5%	467	Ni	6.7
10%	510	Mo	1.6
20%	570	Cu	< 0.1
30%	621	Ca	3.3
40%	670	Mg	1.6
50%	712	Na	17
60%	768	Al	0.77
70%	820	V	0.42
80%	881		
87%	953		
% Over	87.0		
% Bottoms	13.0		

Conradson Carbon, wt-%	1.36
Kjeldahl N, wt-%	1.513
Br. No., Wet Iced	23.6
ASTM Ash, wt-%	0.014
C_7 Insol., wt-%	0.34
H_2O, wt-%	0.05
Arsenic, ppm	27.5
Sulfur, wt-%	0.64
Total O, wt-%	0.65
Carbon, wt-%	84.99
Hydrogen, wt-%	12.27
Diene Value	31.6
Potential Gum, mg/100 mL	15.5

TABLE 80. HEAVY CHARGE STOCK INSPECTION

Charge Stock: Raw Occidental Shale Oil

UOP No.:   49-1098

		General Processing	
Gravity, °API at 60°F	23.3	Sulfur (LECO), wt-%	0.64
Specific Gravity	0.9141	Nitrogen (Total), wt-% (Kj)	1.4
Distillation (D-1160), °F		Total Chloride, ppm	< 1.0
IBP	402	BS and W, vol-%	0.2
5%	489	Conradson Carbon, wt-%	1.17
10%	521	Ash, wt-%	0.00
30%	602	Heptane Insoluble, wt-%	0.43
50%	707		
70%	810	Metal by AAS (ppm)	
90%	946	V	< 2
95%	-	Ni	11
EP	991	Fe	72
% Over	94		
		Total Solids, wt-%	0.17
Pour Point, °F	+ 75		
		C_5 Insoluble, wt-%	1.00

Micro C/H/O/N, Wt-%		Sim. Distillation, °F	
		IBP	325
Carbon	84.5	5%	428
Hydrogen	12.2	30%	577
Nitrogen	1.4	60%	711
		90%	880
		95%	933
Diene	32.9	EP	1017
H_2O, wt-%	0.2		
Potential Gum	15.5	Metal by Emission, ppm	
Total O, wt-%	4.6	Fe	31.0
		Ni	5.6
		V	0.42
		Pb	< 0.1
		Cu	0.07
		Na	11.0
		Mo	1.8

TABLE 81. DESALTED PARAHO SHALE OIL

UOP No. 39-706A

API Gr. at 60°F	19.8	
Sp. Gr. at 60°F	0.9352	

Metal by Emission, ppm	
Fe	38
Ni	2.2
V	0.34
Pb	< 0.1
Cu	0.18
Na	1.1
Mo	0.12

Distillation, °F	
IBP	408
5%	471
10%	525
20%	605
30%	670
40%	730
50%	789
60%	840
70%	889
80%	942
90%	1005

% Over	90.0
% Bottoms	10.0
Conradson Carbon, wt-%	2.71
Kjeldahl N, wt-%	2.186
Br. No., Wet Iced	34.7
ASTM Ash, wt-%	0.007
C_7 Insol., wt-%	0.47
H_2O, wt-%	< 0.01
Arsenic, ppm	19
Sulfur, wt-%	0.70
Total O, wt-%	1.58

TABLE 82. HEAVY CHARGE STOCK INSPECTION

Charge Stock: Raw Paraho Shale Oil

UOP No.:   39-706

		General Processing	
Gravity, °API at 60°F	19.9	Sulfur (LECO), wt-%	0.65
Specific Gravity	0.9346	Nitrogen (Total), wt-%	2.1
Distillation (D-1160), °F		Total Chloride, ppm	< 1.0
IBP	407	BS and W, vol-%	< 0.1
5%	498	Conradson Carbon, wt-%	2.6
10%	533	Ash, wt-%	0.012
30%	683	Heptane Insoluble, wt-%	0.71
50%	802		
70%	904	Metal by AAS (ppm)	
90%	1083	V	< 2
95%	-	Ni	3
% Over	90	Fe	68
Pour Point, °F	+ 85	Total Solids, wt-%	0.04
		C_5 Insoluble, wt-%	1.39
Micro C/H/O/N, Wt-%		Sim. Distillation,	
		IBP	301
Carbon	84.6	5%	415
Hydrogen	11.4	30%	607
		60%	778
		90%	945
		95%	982
		EP	1065
		Metal by Emission, ppm	
		Fe	38
		Ni	2
		V	0.1
		Pb	< 0.1
		Cu	0.05
		Na	3.8
		Mo	< 0.1

TABLE 83. DESALTED LIGHT BERRI CRUDE

UOP No. 109-3422A

API Gr. at 60°F	35.4
Sp. Gr. at 60°F	0.8478
Bubble Pt. at 500 psig, °F	720
Distillation, °F	
IBP	127
5%	225
10%	274
20%	349
30%	420
40%	500
50%	589
60%	682
70%	782
80%	901
90%	1083
% Over	90.0
% Bottoms	10.0
Br. No.	1.2
ASTM Ash, wt-%	0.001
C_7 Insol., wt-%	0.55
H_2O, ppm	323*
Sulfur, wt-%	1.34
Total O, ppm	851
Carbon, wt-%	85.74
Hydrogen, wt-%	12.96
Nitrogen, ppm	810
AAS, ppm	
Cu	< 0.1
Fe	0.55
Ni	2.4

* Interferences

TABLE 84. RAW LIGHT BERRI PETROLEUM

	Whole Crude Oil	Kerosine
UOP Number	109-3424A	109-3423A
API Gr. at 60°F	35.4	46.3
Sp. Gr. at 60°F	0.8478	0.7958
Bubble Pt. at 500 psig, °F	720	815
Distillation, °F		
IBP	126	329
5%	228	403
10%	275	-
20%	355	-
30%	426	-
40%	509	-
50%	594	-
60%	690	
70%	798	-
80%	909	-
89%	1060	
EP	-	528
% Over	89.0	100.00
% Bottoms	11.0	-
ASTM Ash, wt-%	< 0.001	-
C_7 Insol., wt-%	0.53	-
Emission, ppm		
Cu	0.06	< 0.01
Fe	0.33	0.07
Ni	2.1	0.02
V	8.2	0.05
Pb	< 0.5	< 0.03
Na	< 0.5	0.19
Bromine Number	-	< 0.5
Total Sulfur, wt-%	-	0.18
Total Nitrogen, ppm	-	2.4
Carbon, wt-%	-	85.6
Hydrogen, wt-%	-	14.0
Oxygen, ppm	-	877
H_2O, ppm	-	239
Peroxide Number	-	0
Carbonyl Number	-	1.4
Existing Gum, mg/100 mL	-	1.7
Potential Gum, mg/100 mL	-	0.7

TABLE 85. TYPICAL HEAT TRANSFER EQUIPMENT CONDITIONS

	Low Temperature Heat Exchanger, °F	High Temperature Heat Exchanger, °F	Fired Heater, °F
Shell In	200	450	680
Temp. Differential[a]	40	70	160
Wall[b]	240	520	840
Tube Out	280	580	
Shell Out	450	680	720
Temp. Differential	30	70	140
Wall	480	730	860
Tube In	500	780	

(a) Temperature differential between shell and wall

(b) Approximate wall temperature is the mean between the shell and tube fluid temperatures

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TABLE 86. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF LIGHT BERRI KEROSINE FLOW RATE 400 ML/MIN., FLUID TEMP. = 92°C

	Probe Temperature, °C				h		dR_F/dt
Period	Original	Final	Mean	$1/T^1$ x 10^{-4}	Original	Final	x 10^{-5}
			Run No. 1				
4	146.7	147.6	147.2	23.80	359.2	360.4	0
5	157.4	158.3	157.7	23.22	345.2	342.2	1.12
6	180.8	181.3	180.2	22.06	320.2	317.2	0
7	204.3	203.4	203.1	21.00	303.3	306.7	0
8	226.5	228.4	227.6	19.98	297.8	292.8	1.28*
9	253.3	256.9	254.9	18.94	283.4	275.6	2.62*
10	311.5	319.9	316.0	16.98	264.4	251.1	5.05*
11	377.8	425.4	401.0	14.84	243.0	191.2	55.7
12	432.7	462.1	447.0	13.89	185.0	162.0	38.2
13	466.1	494.2	480.0	13.28	158.9	141.8	37.85
14	499.7	515.7	508.0	12.81	138.5	129.9	39.9
			Run No. 2				
6	196.9	196.9	197.3	21.23	263.9	266.3	0
7	220.1	220.2	220.4	20.28	261.7	263.0	0
8	235.7	237.2	237.2	19.60	278.6	272.5	0.1
9	271.0	274.6	273.1	18.31	257.0	249.2	1.6*
10	330.2	337.1	333.6	16.48	244.0	234.5	5.9*
11	424.9	469.2	447.0	13.89	200.3	164.0	39.5

1) $1/T = 1/°Kelvin$

* Arrhenius plot made with these data

TABLE 87. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF LIGHT BERRI DESALTED CRUDE OIL
FLOW RATE 400 ML/MIN., FLUID TEMP. = 92°C

	Probe Temperature, °C				h		dR_F/dt
Period	Original	Final	Mean	$1/K \times 10^{-4}$	Original	Final	$\times 10^{-5}$

Run No. 3

Period	Original	Final	Mean	$1/K \times 10^{-4}$	Original	Final	$\times 10^{-5}$
6	174.2	172.6	173.1	22.42	241.7	248.5	0
7	183.5	184.9	184.5	21.86	248.0	243.0	1.42
8	209.0	209.5	209.6	20.72	244.4	242.9	0
4B	231.8	232.9	232.4	19.79	244.5	242.4	2.13*
5B	263.1	263.2	261.9	18.70	234.4	234.0	3.50*
6B	292.7	298.5	296.0	17.58	229.8	220.8	5.47*
7B	331.7	346.6	339.4	16.33	215.0	196.7	9.90*
8B	381.4	396.2	388.7	15.11	192.5	178.4	9.92*
9B	437.5	457.6	447.5	13.88	171.7	157.0	14.00
10B	496.8	515.6	506.9	12.82	154.9	143.7	13.80

Run No. 4

Period	Original	Final	Mean	$1/K \times 10^{-4}$	Original	Final	$\times 10^{-5}$
5	172.5	173.3	173.1	22.41	280.2	276.3	1.41
6	196.4	195.3	195.8	21.33	273.8	273.6	0
7	219.3	220.8	219.3	20.31	267.8	264.6	1.36*
8	247.2	247.8	247.7	19.20	259.1	255.5	1.42*

* Arrhenius plot made with these data

TABLE 88. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF
LIGHT BERRI DESALTED CRUDE OIL.
FLUID TEMP. = 90°C, FLUID FLOW = 100 ML/MIN

| Period | Probe Temperature, °C | | | | h | | dR_F/dt |
	Original	Final	Mean	$1/K$ x 10^{-4}	Original	Final	x 10^{-5}
			Run No. 5				
6							
7	180.5	181.1	180.6	22.05	230.2	229.3	0.79
8	204.7	205.2	204.9	20.96	231.5	228.3	1.75*
9	228.4	228.4	228.2	19.96	228.9	228.9	0.82
10	254.8	257.4	256.0	18.90	228.0	222.9	2.50*
11	284.6	289.5	289.1	17.89	224.4	214.6	6.28*
12	318.7	331.0	324.8	16.73	212.3	197.1	12.10
13	364.7	379.3	372.9	15.48	190.7	176.8	14.11*
14	414.0	438.6	426.6	14.29	174.0	155.1	22.65*
15	478.0	496.5	486.2	13.17	151.6	140.6	16.67
16	537.1	545.9	540.4	12.27	138.0	133.6	9.64

* Arrhenius Plot made with these Data

TABLE 89. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES
OF LIGHT BERRI RAW CRUDE OIL
FLUID TEMP = 92°C

| | Probe Temperature, °C | | | | h | | dR_F/dt |
Period	Original	Final	Mean	$1/K$ $\times 10^{-4}$	Original	Final	$\times 10^{-5}$
6	174.0	175.5	174.7	22.34	221.4	223.1	NIL
7	186.8	187.6	187.2	21.73	224.3	222.7	0.8*
8	211.8	212.3	212.2	20.61	225.3	224.7	0.7*
9	236.6	236.4	236.4	19.63	225.2	224.5	NIL
10	263.0	262.2	262.6	18.67	224.1	225.8	NIL
11	290.1	290.7	290.0	17.76	224.7	224.5	0.93*
12	319.3	322.2	320.9	16.86	223.1	218.6	3.04*
13	352.6	362.0	359.6	15.80	216.1	204.8	6.15*
14	397.0	404.1	400.8	14.84	199.4	192.3	6.78*
15	442.5	446.9	444.4	13.94	185.9	182.0	3.12
16	484.9	487.1	485.1	13.19	177.8	176.2	3.08

* Arrhenius plot made with these data

TABLE 90. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF RAW LIGHT BERRI CRUDE INHIBITED WITH 100 PPM OF ANTIFOULANT

| Period | Probe Temperature, °C | | | | h | | dR_F/dt |
	Original	Final	Mean	$1/K$ $\times 10^{-4}$	Original	Final	$\times 10^{-5}$
6	182.1	183.4	182.4	21.96	210.0	216.7	NIL
7	194.3	194.3	194.4	21.40	216.5	217.6	1
8	220.0	220.5	220.5	20.26	220	218.8	1
9	246.1	245.4	245.7	19.27	219.0	220.7	NIL
10	274.3	275.9	274.6	18.26	219.4	217.0	2.60*
11	306.5	310.2	308.7	17.19	214.2	209.6	3.42*
12	342.1	345.9	344.5	16.19	207.2	202.7	3.02*
13	380.0	385.8	383.1	14.24	199.4	193.8	5.05*
14	423.9	436.8	431.9	14.19	188.5	177.6	8.58*
15	472.8	481.5	476.9	13.33	176.0	170.1	7.17
16	519.9	530.6	523.2	12.51	167.4	160.0	7.89

* Arrhenius plot made with these data

TABLE 91. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF
OCCIDENTAL DESALTED SHALE OIL
FLUID TEMP. = 92°C, FLOW = 100 ML/MIN.

	Probe Temperature, °C				h		dR_F/dt
Period	Original	Final	Mean	$1/K$ $\times 10^{-4}$	Original	Final	$\times 10^{-5}$

Run No. 7

6	173.2	174.3	173.6	22.4	225.4	227.6	0.59
7	185.6	187.1	186.4	21.8	228.5	223.8	2.79*
8	211.8	215.0	213.5	20.6	226.9	218.2	4.86*
9	239.6	243.9	242.0	19.4	217.7	209.8	6.08*
10	273.4	279.7	276.3	18.2	208.1	197.7	7.72*
11	310.9	320.0	315.9	17.0	195.3	184.0	11.22*
12	356.2	370.0	363.4	15.7	179.4	166.2	14.90*
13	411.3	450.9	430.8	14.2	169.8	133.1	46.8
14	494.4	520.1	506.2	12.8	129.9	117.5	30.82
15	568.0	591.5	578.4	11.7	114.8	105.7	27.7
16	641.0	668.2	654.6	10.8	103.7	95.5	28.1

Run No. 9

6	175.0	175.2	174.7	22.34	219.8	222.7	0.02
7	185.8	186.7	186.3	21.77	227.0	224.9	0.19
8	210.9	212.7	211.8	20.63	227.3	223.6	3.61*
9	237.0	240.2	238.4	19.55	223.2	217.2	5.62*
10	268.2	272.9	270.6	18.40	216.1	208.0	7.22*
11	303.4	310.3	306.6	17.25	205.2	196.2	8.17*
12	343.1	353.7	348.4	16.09	192.9	181.2	11.45*
13	388.3	401.9	395.3	14.96	178.4	166.4	13.03*
14	441.3	469.0	456.9	13.70	162.0	143.2	28.54
15	511.0	533.1	521.0	12.59	140.3	129.1	25.40
16	578.6	592.3	583.0	11.68	125.9	120.1	17.75

* Arrhenius plot made with these data

TABLE 92. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF
RAW OCCIDENTAL SHALE OIL
FLUID TEMP = 92°C.

| Period | Temperature | | | | h | | dR_F/dt |
	Original	Final	Mean	$1/K$ $\times 10^{-4}$	Original	Final	$\times 10^{-5}$
6	170.2	173.1	171.4	22.50	223.6	228.8	1.15
7	184.0	187.7	185.9	21.79	230.1	219.8	6.35*
8	212.5	218.9	215.9	20.45	220.2	206.2	10.96*
9	245.6	254.0	250.0	19.12	202.5	187.4	13.80*
10	285.2	295.9	290.9	17.73	183.5	171.5	13.78*
11	331.3	345.3	338.3	16.36	166.9	153.0	17.64*
12	384.6	404.7	393.3	15.01	148.5	133.8	26.93*
13	451.6	482.6	468.0	13.50	128.2	112.1	36.34*
14	529.7	564.1	539	12.32	109.2	96.4	44.82
15	614.8	672.2	644.5	10.90	94.3	79.1	67.35
16	729.1	801.4	766.7	9.61	77.3	64.2	95.25

* Arrhenius plot made with these data

TABLE 93. EFFECT OF TEMPERATURE ON h AND dR_F/dt VALUES OF
DESALTED PARAHO SHALE OIL
FLUID TEMP = 92°C

	Temperature, °C				h		dR_F/dt
Period	Original	Final	Mean	$1/K \times 10^{-4}$	Original	Final	$\times 10^{-5}$
6	167.6	173.5	170.7	22.54	241.2	229.7	8.64*
7	185.1	189.7	187.7	21.69	229.4	217.9	7.63*
8	216.3	221.4	218.8	20.33	215.4	203.6	9.72*
9	247.9	251.9	250.0	19.12	201.6	194.8	6.04*
10	283.5	285.6	284.1	17.95	191.3	187.7	3.67*
11	319.3	320.7	319.8	16.87	183.5	181.9	2.27*
12	355.6	362.3	355.9	15.90	177.7	173.8	6.99*
13	399.0	404.6	402.5	14.80	169.0	163.3	8.84*
14	447.4	457.4	450.1	13.83	157.0	149.8	14.79*
15	499.3	520.1	511.8	12.74	146.8	134.3	23.25
16	558.6	582.8	570.6	11.85	133.5	123.0	22.07

* Arrhenius plot made with these data

TABLE 94. EFFECT OF TEMPERATURE AND ANTIFOULANT INHIBITOR
ON h AND dR_F/dt VALUES OF
DESALTED PARAHO SHALE OIL

| Period | Temperature, °C | | | | h | | dR_F/dt |
	Original	Final	Mean	$1/K$ $\times 10^{-4}$	Original	Final	$\times 10^{-5}$
6	171.2	172.7	171.7	22.49	235.6	234.8	NIL
7	183.9	186.5	185.3	21.82	236.9	229.5	3.91*
8	210.5	213.5	212.1	20.61	231.0	224.1	3.75*
9	237.5	239.9	238.7	19.54	224.1	219.5	2.90*
10	268.5	270.2	269.1	18.45	217.8	213.8	2.95*
11	300.4	303.6	301.5	17.41	210.7	206.8	3.44*
12	335.3	338.7	336.5	16.41	203.1	198.7	4.25*
13	372.1	377.6	374.8	15.44	195.7	189.8	5.55*
14	414.8	432.6	422.8	14.37	184.1	169.5	17.31*
15	472.6	496.8	485.5	13.18	164.6	148.9	21.69
16	535.4	559.2	547.0	12.20	147.5	135.3	21.23

* Arrhenius plot made with these data

TABLE 95. ACTIVATION ENERGIES OF LIGHT BERRI, OCCIDENTAL AND PARAHO PRODUCTS

Product	Activation Energy, cal
Raw Light Berri Crude Oil	-6897
Raw Light Berri Crude Oil + Antifoulant	-5472
Desalted Light Berri Crude Oil	-8176
Light Berri Kerosine	-8566
Raw Occidental Shale Oil	-3758
Desalted Occidental Shale Oil	-4626
Desalted Paraho Shale Oil	
167 to 315°C	+2846
315 to 475°C	-11338
Desalted Paraho Shale Oil with Inhibitor	
167 to 315°C	+915
315 to 475°C	-10168
Light Berri Desalted Crude Oil	
1) at 400 mL/min.	-8753
2) at 100 mL/min.	-8176

TABLE 96. EFFECT OF TEMPERATURE AND FLOW RATE ON dR_F/dt OF
DESALTED LIGHT BERRI AND DESALTED OCCIDENTAL SHALE OIL

Temperature		$dR_F/dt \times 10^{-5}$ Light Berri Desalted Crude Oil	
°C	°F	400	100 ml/min
200	392	1.05	1.32
225	437	1.88	2.27
250	482	2.86	3.36
275	527	4.20	4.80
300	572	5.97	6.67
325	617	8.23	9.01
350	662	11.05	11.86
375	707	14.51	15.30
400	752	18.64	19.35

TABLE 97. EFFECT OF WIRE TEMPERATURE ON dR_F/dt OF
LIGHT BERRI CRUDE OIL AND KEROSINE
100 ml/min.

Wire Temperature		$dR_F/dt \times 10^{-5}$			
°C	°F	Crude Oil	Inhibited Crude Oil	Desalted Crude Oil	Kerosine
200	392	0.60	1.00	1.32	0.68
225	437	0.96	1.42	2.27	0.85
250	482	1.33	1.85	3.36	1.81
275	527	1.80	2.34	4.80	2.63
300	572	2.38	2.92	6.67	3.70
325	617	3.06	3.57	9.01	5.08
350	662	3.86	4.29	11.86	6.78
375	707	4.79	5.09	15.30	8.85
400	752	5.83	5.95	19.34	11.31
Activation Energy		-7061	-5472	-8176	-8566

TABLE 98. EFFECT OF TEMPERATURE ON dR_F/dt OF
OCCIDENTAL AND PARAHO SHALE OILS

Wire Temperature		$dR_F/dt \times 10^{-5}$			
		Occidental Shale Oil		Paraho Desalted Shale Oil	
°C	°F	Crude	Desalted	Original	Inhibited
200	392	7.89	3.49	9.0	3.0
225	437	8.67	4.75	9.0	3.0
250	482	11.94	5.94	6.0	3.0
275	527	13.99	7.27	4.2	3.0
300	572	16.17	8.75	3.0	3.0
325	617	18.47	10.37	2.3	4.0
350	662	20.82	12.12	4.5	5.7
375	707	23.34	14.00	6.4	7.8
400	752	25.89	15.99	8.9	10.5
Activation Energy		-3616	-4629	-	-

TABLE 99. EFFECT OF TEMPERATURE, TIME OF REACTION, AND NEW TEST PROBE
ON h AND dR_F/dt VALUES OF DESALTED OCCIDENTAL SHALE OIL
FLUID TEMP = 92°C

| Time, Hrs. | Wire Temperature | | | | h | | dR_F/dt |
	Original	Final	Mean	$1/K \times 10^{-4}$	Original	Final	$\times 10^{-5}$
			Run No. 13				
0-3	250.6	255.5	248.8	19.10	254.5	248.8	4.81
6-9	262.0	268.3	265.0	18.58	234.5	223.1	7.45
12-15	274.1	279.3	276.9	18.18	213.0	204.2	6.57
18-21	283.8	288.1	286.0	17.89	197.2	191.2	5.20
24-27	293.5	297.4	295.0	17.61	184.8	179.1	6.26
30-33	301.5	306.6	303.9	17.33	173.8	168.6	5.73
36-39	309.9	314.1	312.0	17.09	164.2	159.4	6.04
42-45	318.5	321.8	319.9	16.87	155.0	152.0	5.06
			Run No. 14				
0-3.9	279.0	285.6	281.6	18.03	243.9	235.6	5.76
6-9	290.8	300.4	295.6	17.59	226.1	211.9	10.51
9-12	295.5	307.7	301.1	17.41	218.9	201.1	10.84
15-18	321.5	330.5	325.8	16.72	183.2	173.4	10.79
21-24	341.4	349.7	345.8	16.16	161.5	153.5	9.91
27-30	357.1	364.1	361.0	15.78	147.0	141.6	9.12
33-36	371.0	377.7	374.6	15.46	136.3	131.1	9.76
39-40.5	384.7	388.5	386.7	15.16	126.4	123.9	9.10
			Run No. 15				
0-3	311.4	322.8	317.7	16.94	239.1	224.8	10.36
6-9	332.5	344.6	338.7	16.35	211.4	196.5	11.56
12-15	357.8	375.5	365.7	15.66	182.0	164.5	19.12
18-21	407.6	425.6	416.4	14.50	139.8	127.9	23.49
24-27	445.1	462.4	453.4	13.77	116.5	108.2	23.29
30-33	478.8	493.1	485.7	13.18	101.0	95.0	20.81
36-39	504.5	515.2	510.0	12.77	90.8	87.1	15.41

TABLE 100. COMPARISON OF INCREMENTAL INCREASE IN WIRE TEMPERATURES
AND CONSTANT TEMPERATURE PROCEDURES

| | Incremental Increase In Wire Temp | $dR_F/dt \times 10^{-5}$ | | | | |
| | | Constant Temperature | | | | |
		0-3 Hrs.	6-9 Hrs.	12-18 Hrs.	0.42 Hrs.	Ave. 0-18 Hrs.
Fouling Rate ($\times 10^{-5}$)						
At 227°C	4.75	3.27	6.0	2.96	3.44	4.07 ± 1.67
At 282°C	7.40	6.65	8.8	6.85	10.5	7.43 ± 1.19
At 352°C	12.10	13.55	12.92	15.82	32.3	14.10 ± 1.52
Activation Energy	-4629	-7163	-3867	-8332	-11150	
Corr. Coefficient		-0.96	-0.93	-0.99	-0.99	

TABLE 101. EFFECT OF INCREASING AND DECREASING WIRE PROBE TEMPERATURE ON
h AND dR_F/dt OF OCCIDENTAL SHALE OIL

Period	Probe Temperature, °C			$1/K$ x 10^{-4}	h		dR_F/dt x 10^{-5}
	Original	Final	Mean		Original	Final	
				Run No. 8			
6	223.8	226.9	225.6	20.06	242.9	238.1	2.87
7	334.5	356.0	345.4	16.17	232.6	203.5	22.42
8	424.3	476.2	449.8	13.83	194.9	154.4	50.55
9	519.5	585.5	554.5	12.08	149.5	118.1	36.63
10	542.6	569.1	555.7	12.07	119.5	108.5	30.62
11	478.4	490.3	483.7	13.21	112.7	107.4	16.79
12	327.2	328.0	327.9	16.64	108.2	107.6	0.60
				Run No. 10			
6	246.8	245.5	245.3	19.29	208.3	214.0	0
7	362.8	379.0	371.7	15.51	213.8	195.9	13.76
8	449.5	488.9	466.9	13.51	189.8	160.5	35.81
9	526.8	578.2	559.2	12.01	159.2	132.6	27.29
10	536.3	558.9	546.9	12.20	133.7	126.3	23.40
11	471.3	480.0	475.3	13.36	126.7	122.3	10.79
12	325.7	325.5	325.5	16.71	117.9	118.0	0.08

FLOW

ELECTRICAL
INSULATION AND
HIGH PRESSURE
SEALANT

ELECTRICAL
BUSBAR

CONAX
CONNECTORS
FOR 1/16 INCH
THERMOCOUPLE

'A'   'A'

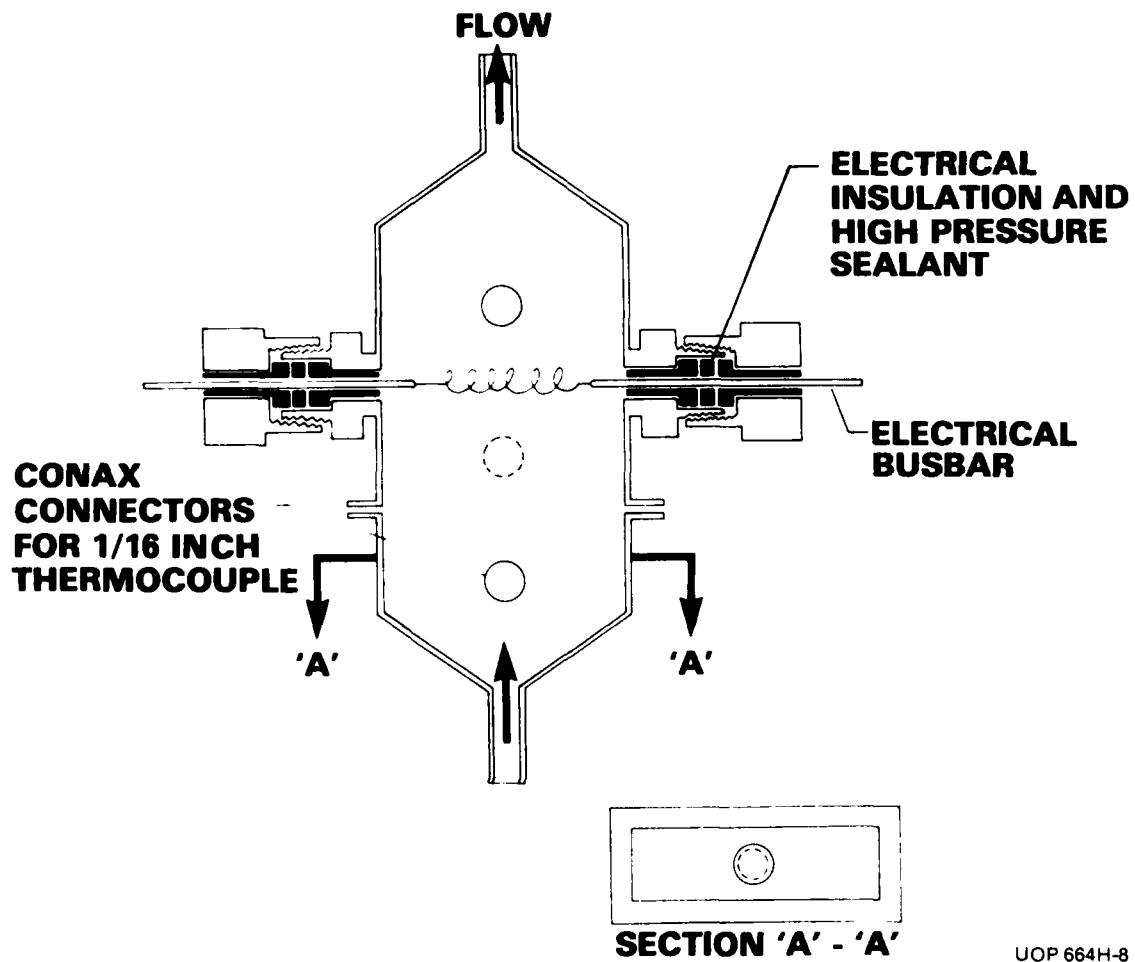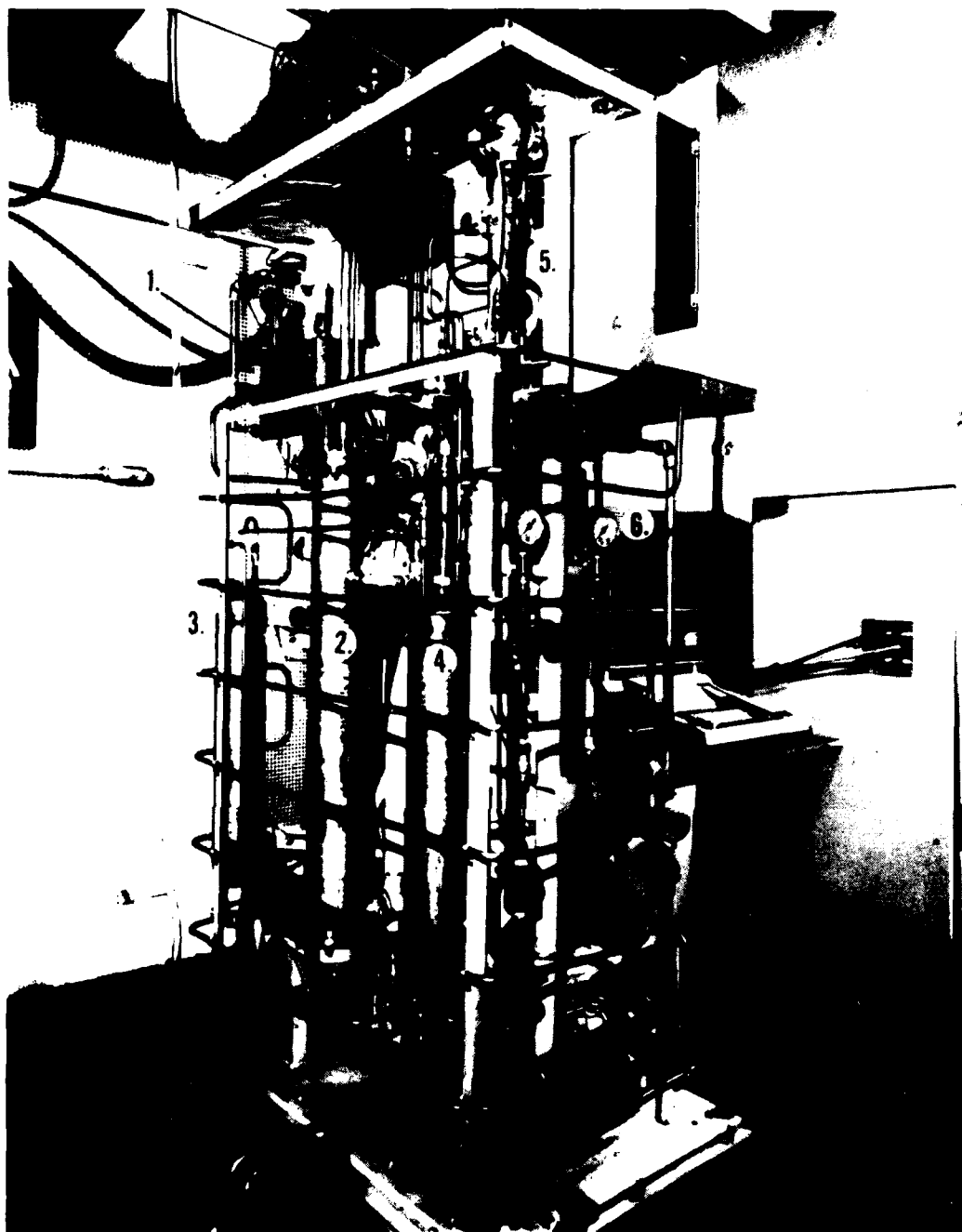**SECTION 'A' - 'A'**   UOP 664H-8

FIGURE 45
UOP FOULING PROBE

UOP 663-34

FIGURE 46

FOULING MONITOR
(FRONT VIEW)
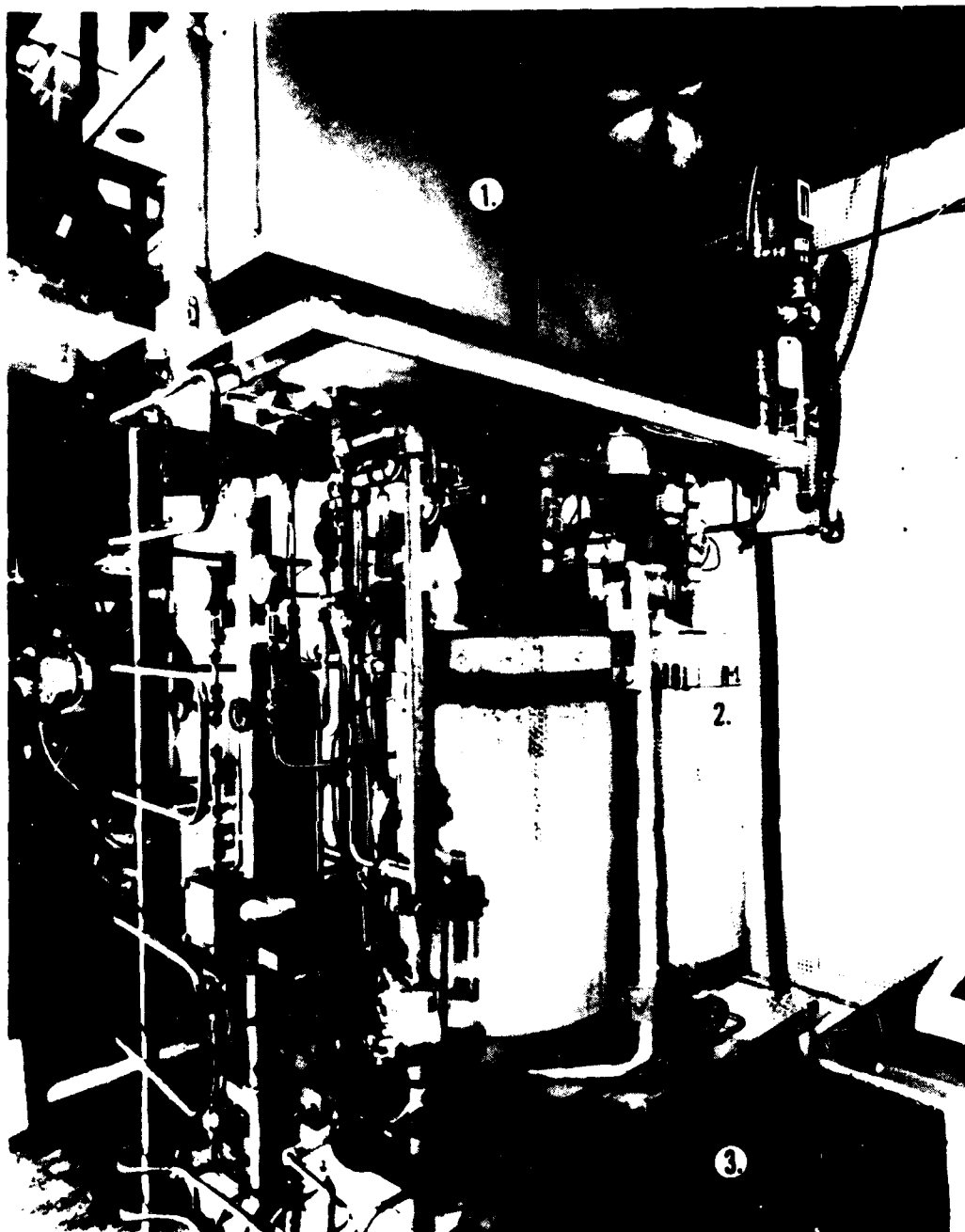
UOP 663-35

FIGURE 47
FOULING MONITOR
(BACK VIEW)
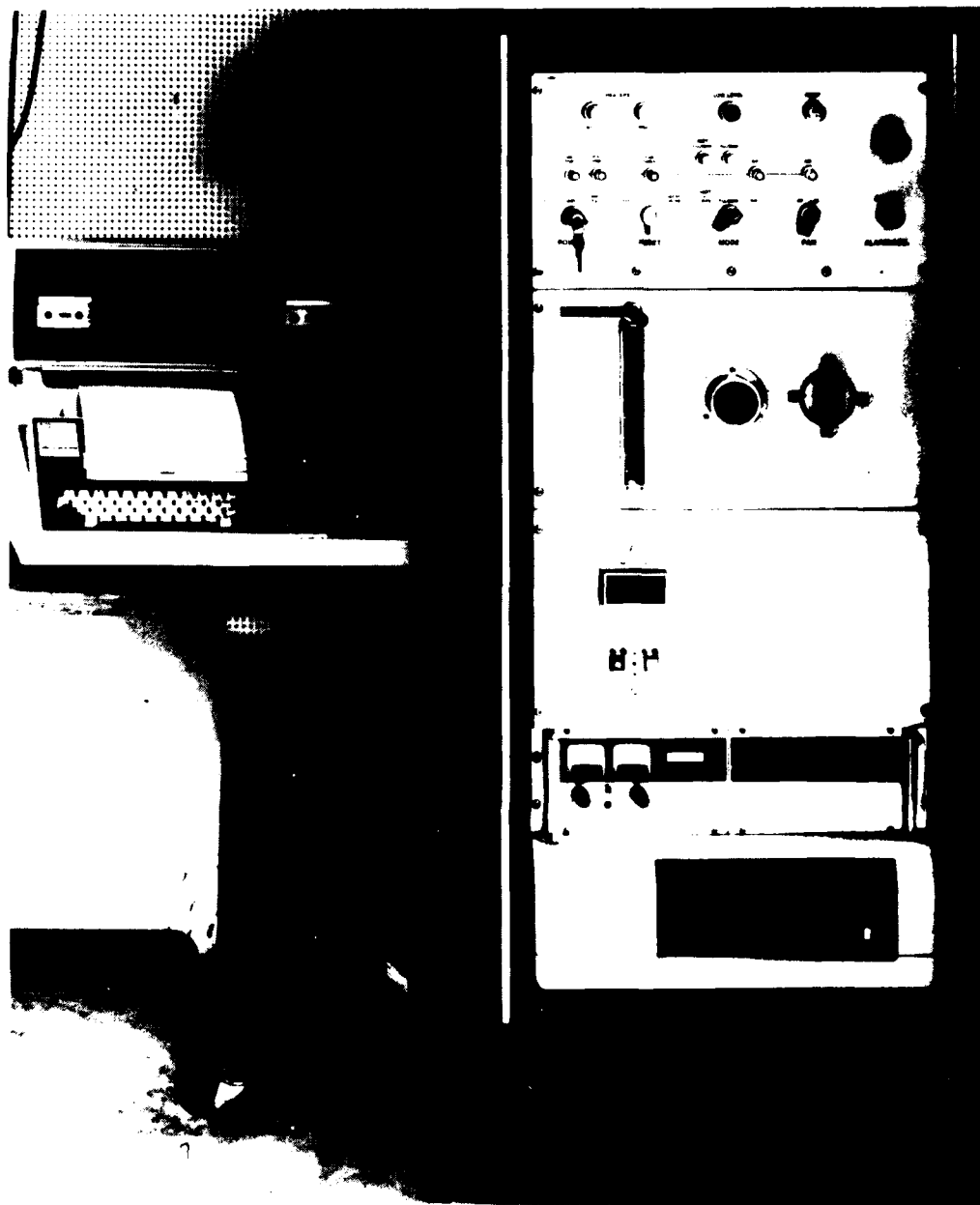
UOP 663-36

FIGURE 48

CONTROL CABINET AND DATA TERMINAL
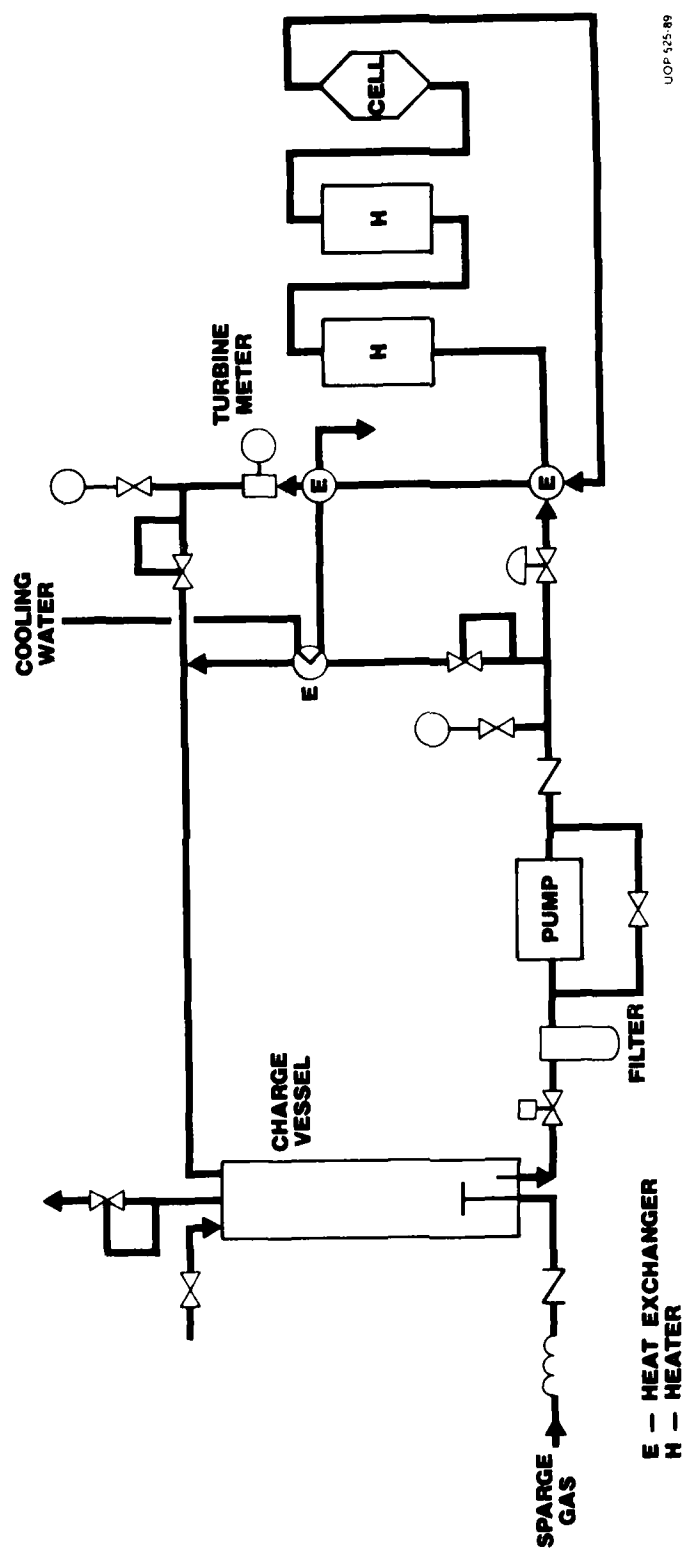
FIGURE 49

UOP MONIREX® FOULING MONITOR

UOP 525-89

E — HEAT EXCHANGER
H — HEATER
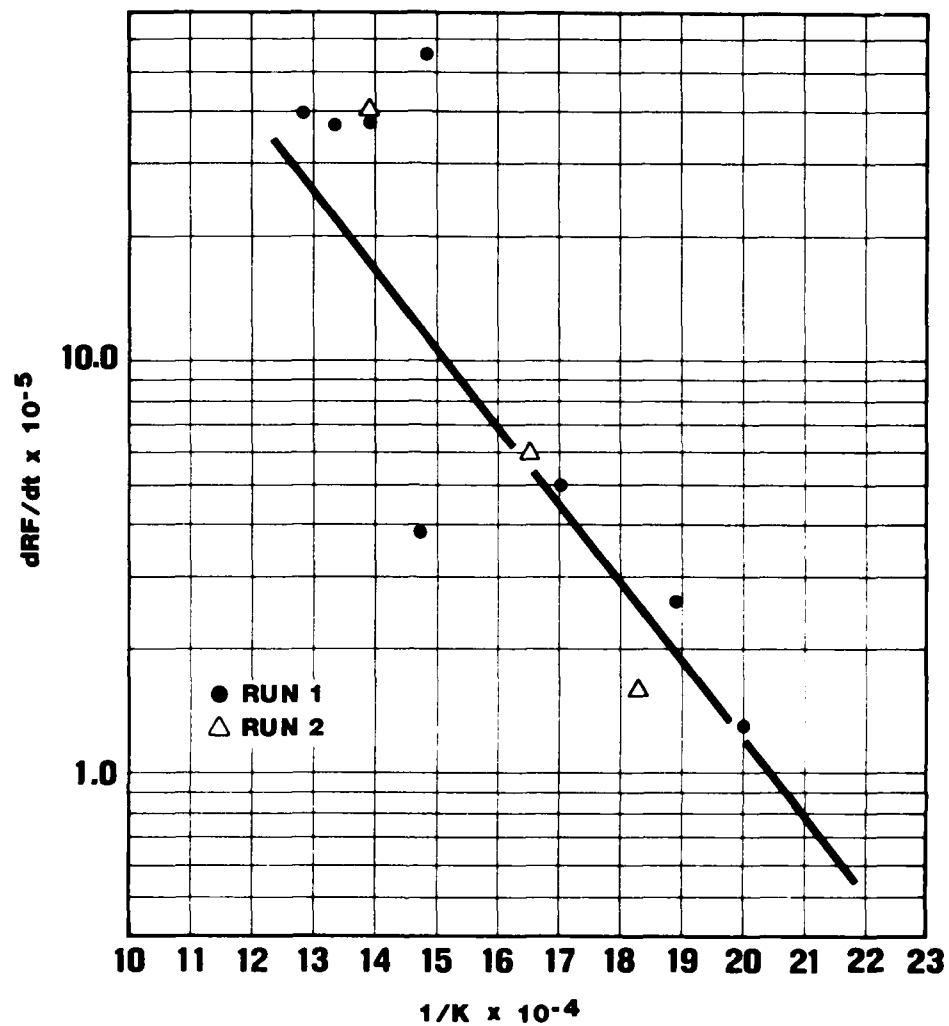
FIGURE 50

EFFECT OF TEMPERATURE ON
dRF/dt OF LIGHT BERRI KEROSINE
(FLOW RATE = 400 ML, FLUID TEMP. = 92°C)
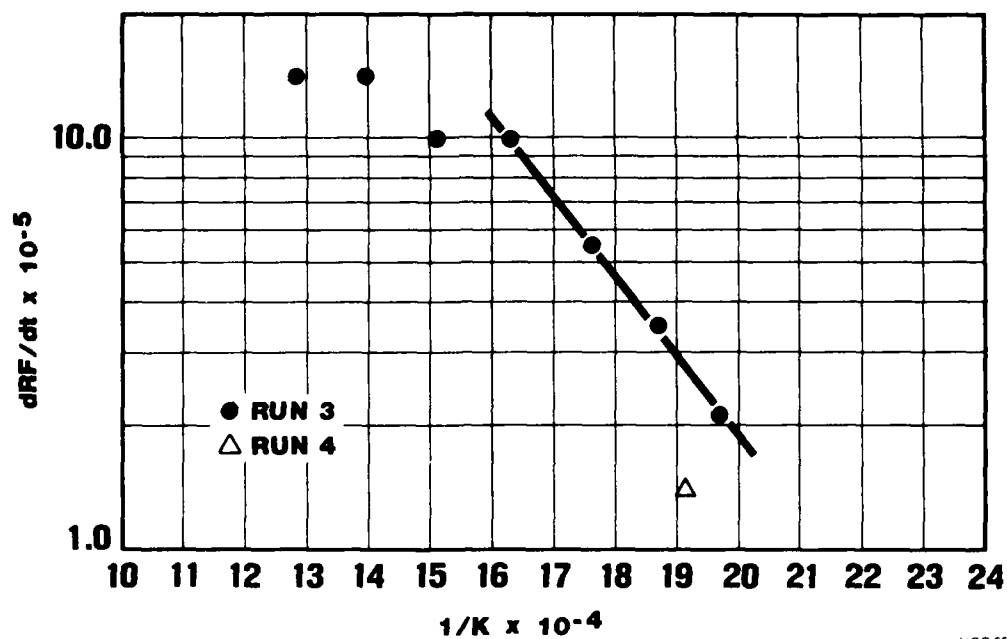
FIGURE 51

EFFECT OF TEMPERATURE ON dRF/dt x 10⁻⁵ OF LIGHT BERRI DESALTED CRUDE

(FLOW RATE = 400 ML/MIN, RUNS 3 AND 4)
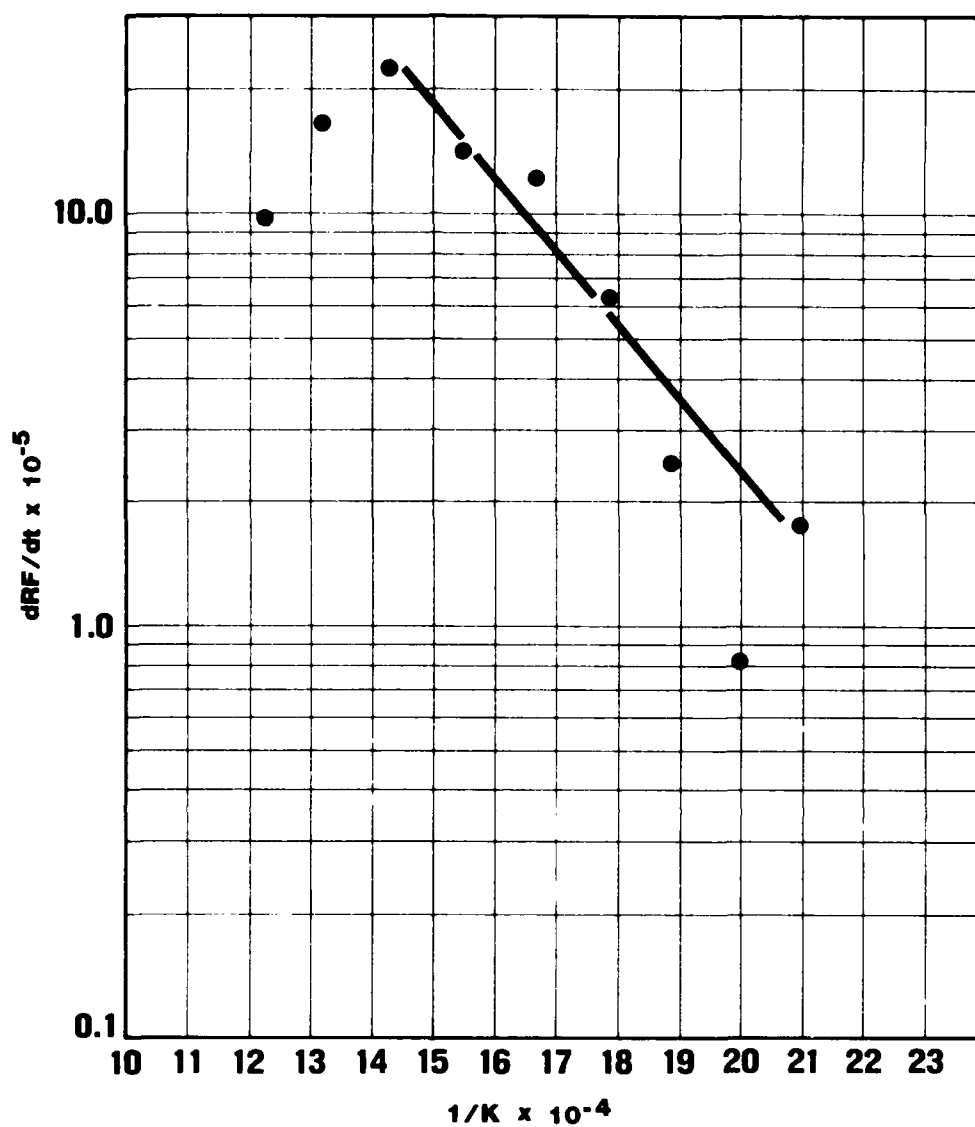
FIGURE 52

**EFFECT OF TEMPERATURE ON
LIGHT BERRI DESALTED CRUDE dRF/dt**

(FLOWRATE = 100 ML/MIN, FLUID TEMP. = 92°C)
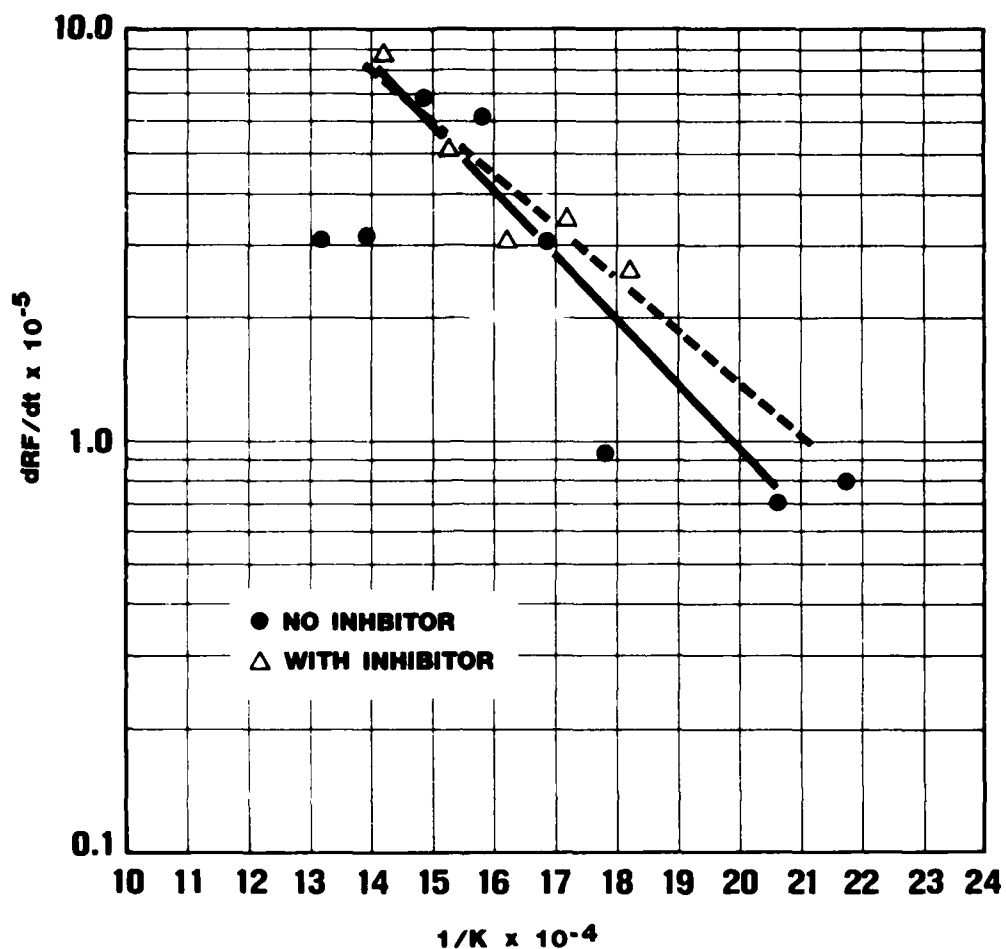
UOP 663-9

FIGURE 53

**EFFECT OF TEMPERATURE AND
ANTIFOULANT INHIBITOR ON
dRF/dt OF RAW LIGHT BERRI CRUDE**
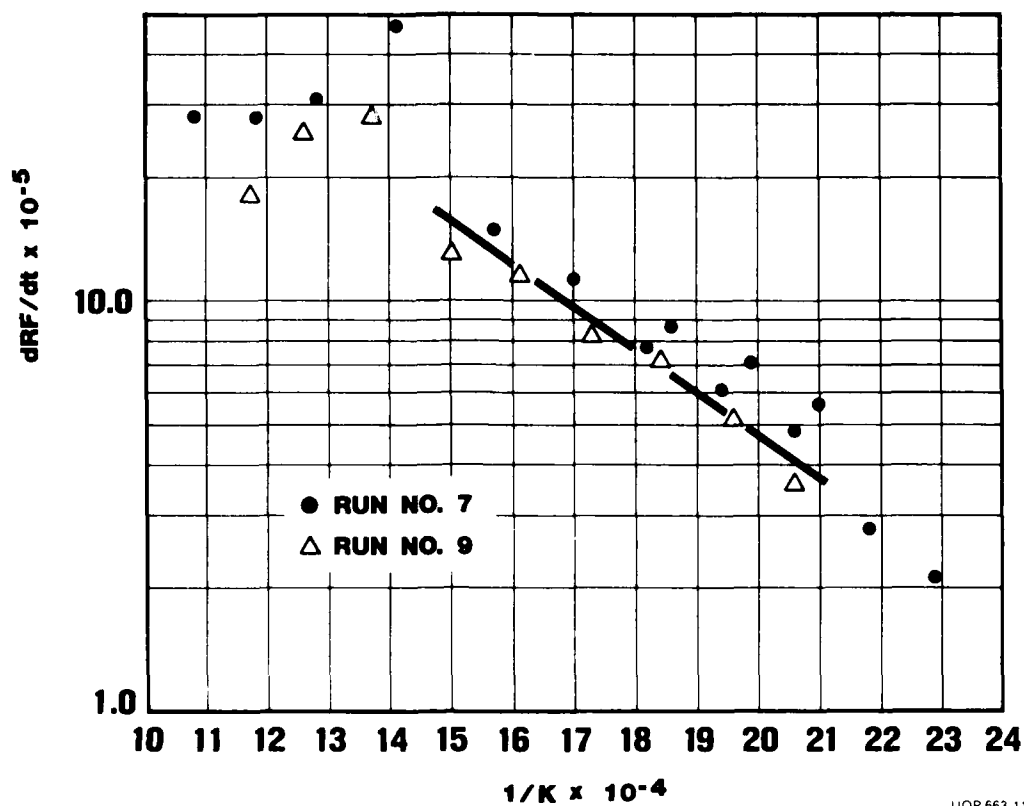
FIGURE 54

**EFFECT OF TEMPERATURE
ON dRF/dt OF DESALTED
OCCIDENTAL SHALE OIL**

(FLOW RATE = 100 ML/MIN, FLUID TEMP. = 92°C)
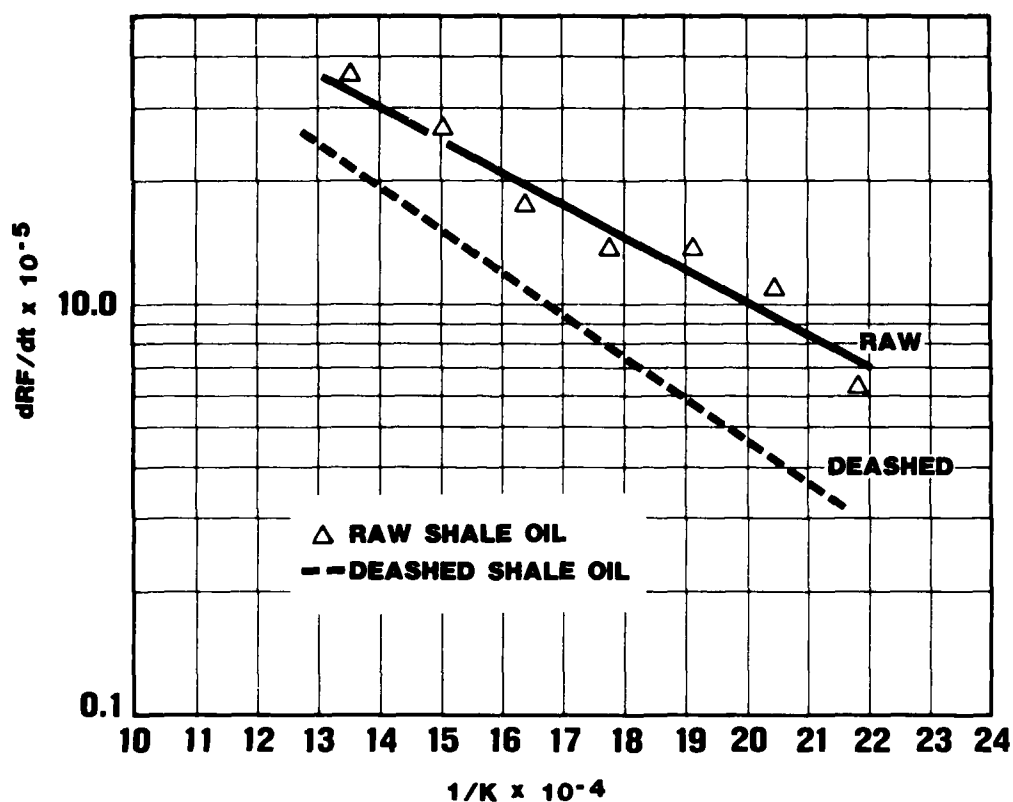
FIGURE 55

**EFFECT OF TEMPERATURE ON
dRF/dt OF OCCIDENTAL SHALE OILS**

(FLOW RATE = 100 ML/MIN, FLUID TEMP. = 92°C)
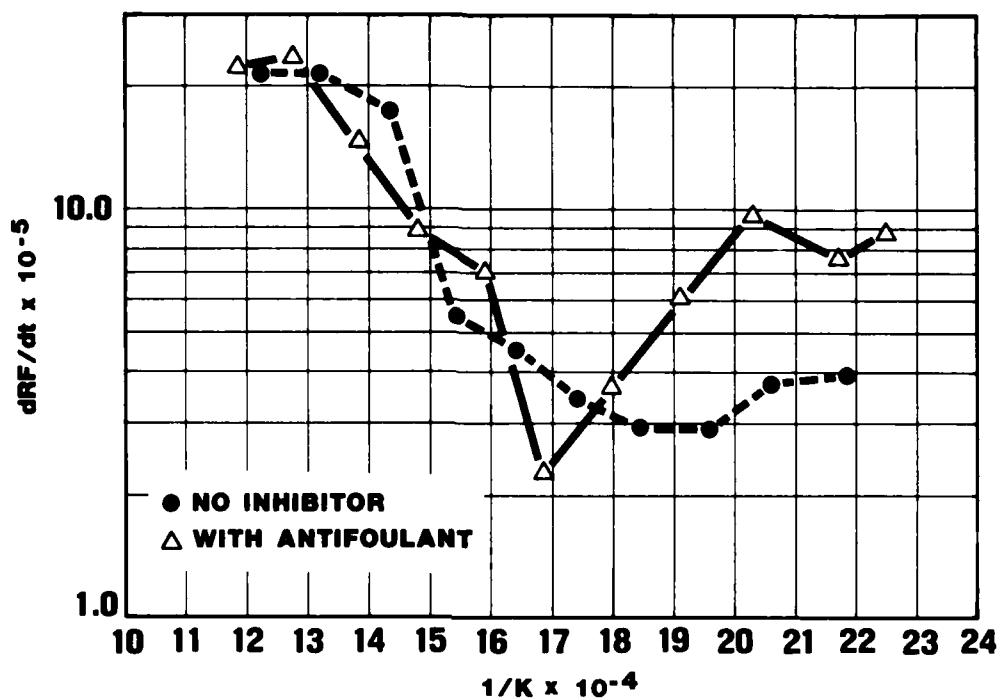
FIGURE 56

EFFECT OF TEMPERATURE AND ANTIFOULANT INHIBITOR ON dRF/dt OF DESALTED PARAHO SHALE OIL
(FLOW RATE = 100 ML/MIN, FLUID TEMP. = 92°C)
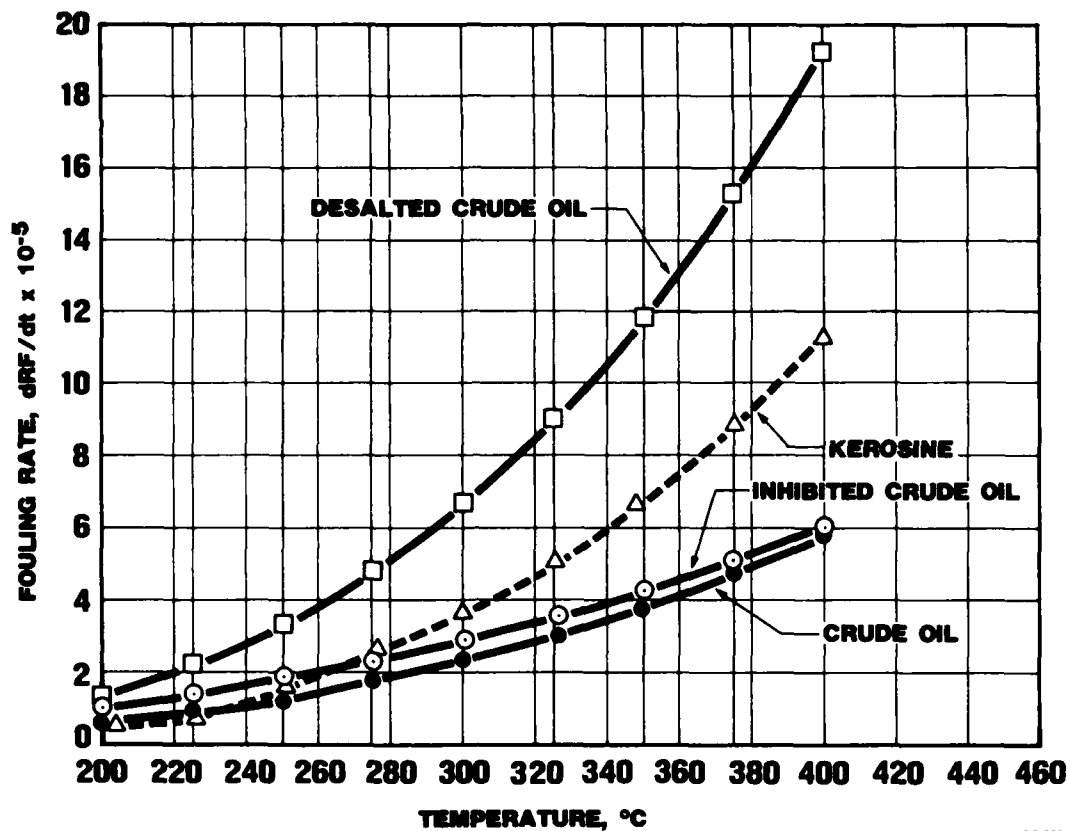
FIGURE 57

**EFFECT OF TEMPERATURE, ON
dRF/dt OF LIGHT BERRI CRUDE,
DESALTED CRUDE PETROLEUM
OILS AND LIGHT BERRI KEROSINE**
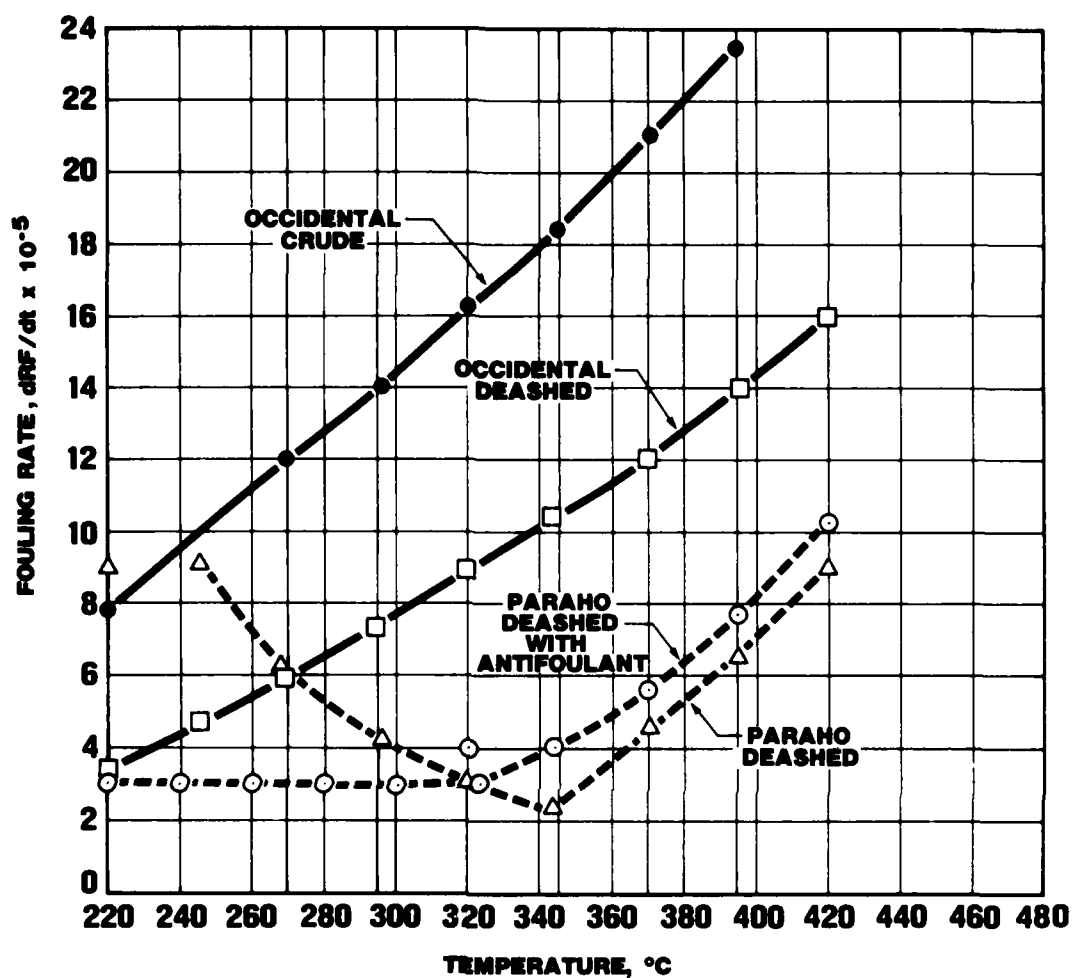
FIGURE 58

**EFFECT OF TEMPERATURE ON
dRF/dt OF OCCIDENTAL CRUDE AND
DESALTED PARAHO SHALE OILS**
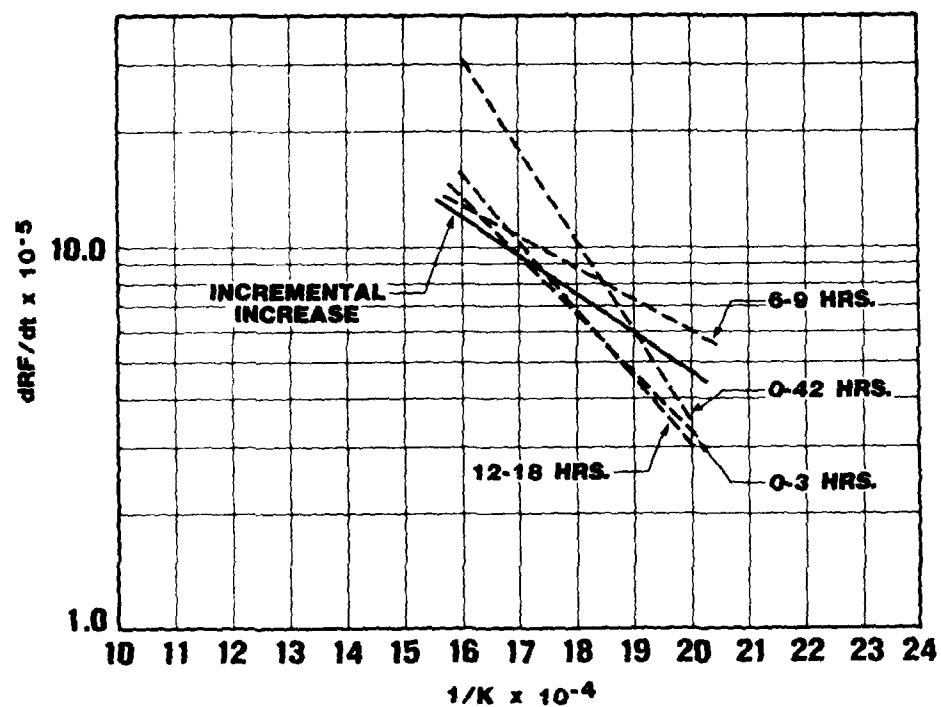
FIGURE 59
**FOULING RATES AT VARIOUS
TEMPERATURES OBTAINED WITH
INCREMENTAL STEP INCREASE AND
CONSTANT TEMPERATURE TECHNIQUES**
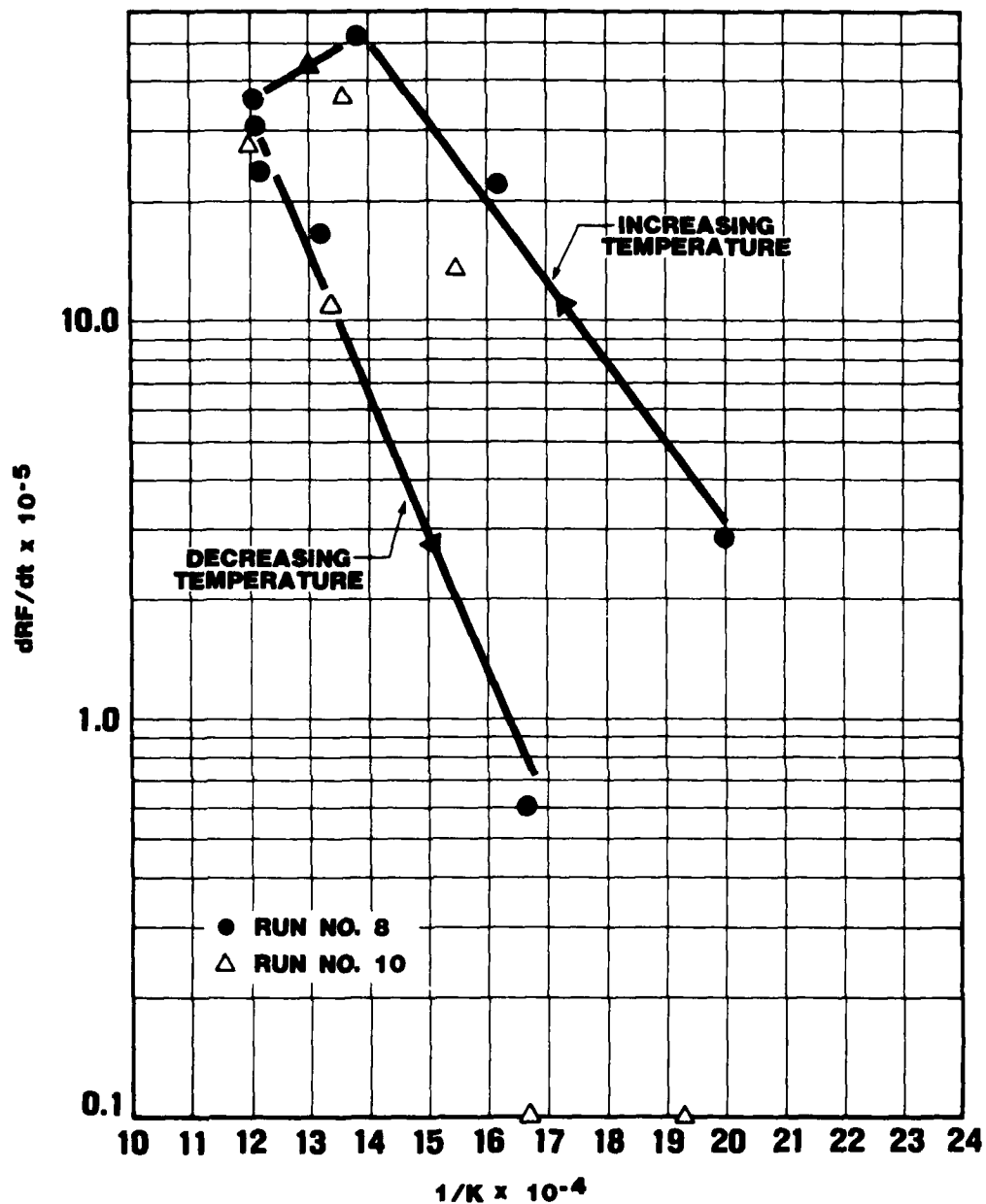
FIGURE 60

EFFECT OF VARYING
WIRE PROBE TEMPERATURE ON dRF/dt
OF DESALTED OCCIDENTAL SHALE OIL

SECTION IV

PROCESS AND ECONOMIC EVALUATION

Basis and Scope of Study

The following is a list of basic information and assumptions used to develop the economic evaluation for the various refining processing schemes to upgrade and refine the Occidental shale oil feedstock into acceptable jet fuels and coproduct fuels. Data generated during the UOP pilot plant operations on the shale oil feedstock were to be incorporated into this evaluation.

A. The refinery design and flow schemes are to be based on processing 100,000 BPSD of 22.9 °API desalted Occidental shale oil. The Occidental shale oil inspection report is as presented in Table 102.

B. The major processing scheme uses two stages of hydrotreating followed by hydrocracking to produce acceptable jet fuel products.

C. Two types of jet fuel products are required.

1. Aviation turbine fuel grade JP-4 covered by Military Specification MIL-T-5624K, dated April 1, 1976.

2. Aviation turbine fuel grade JP-8 covered by Military Specification MIL-T-83133, dated May 5, 1976.

D. Desired coproducts from the processing schemes are:

1. Diesel fuel marine covered by Military Specification MIL-F-16884G, dated March 22, 1978

2. Diesel fuel DF-2 covered by Federal Specification VV-F-800B, dated April 2, 1975

3. Combat gasoline F-46 covered by Military Specification MIL-G-3056D, dated September 29, 1975

4. Fuel oil covered by Military Specification MIL-F-869E, dated September 22, 1965

5. Aviation gasoline 100/130 covered by Military Specification MIL-G-5572E, dated March 20, 1975

6. JP-5 turbine fuel covered by Military Specification MIL-T-5624K, dated April 1, 1976

7. Automotive gasoline, unleaded special grade No. 1 covered by Federal Specification VV-G-1690B, dated July 1, 1978

E. Only commercially proven processes are to be considered in the refinery processing schemes.

F. There are no outside feed streams to be processed with the raw shale oil. Fuels for heating are to be generated by the refinery.

G. The refinery is assumed to be located at a midwest site adjacent to an existing refinery where excess steam and fuels can be credited.

H. Capital Investment

1. Midwest location
2. 100,000 BPSD crude shale oil capacity
3. Second quarter 1980 cost base (capital cost in this study first quarter of 1981)
4. $150,000,000 offsites allowance
5. 100% equity financing
6. Investment timing over a three year construction period.
    a. 25% first year
    b. 50% second year
    c. 25% third year
7. 10% investment tax credit

I. Working Capital

1. 21 days crude storage capacity with 14 day crude inventory
2. 14 days product storage capacity with 7 day product inventory
3. Crude valued at $30.00 per barrel
4. Product valued at $40.00 per barrel (only to determine working capital)
5. Debt financed at 10%

J. Capital Return

1. 15% DCF rate of return on capital
2. 13 years sum of years digits depreciation
3. Zero salvage value

K. Operating Basis

1. 16 year plant operating life
2. 50% operating capacity for 1st year -- 100% capacity thereafter
3. 90% on-stream factor
4. 100,000 BPSD capacity
5. All process fuel/heat requirements shall be generated internally from the original shale oil feed.

L. Operating Cost Basis:

1. Crude Shale Oil - $30.00 per barrel
2. Cooling Water - $0.03 per 1000 gals.
3. Electricity - $0.035 per kWh
4. Operators - $12.00 per hour
5. Helpers - $10.50 per hour
6. Labor Supervision - 25% of direct labor
7. Overhead - 100% of direct labor
8. Federal and State Taxes - 50%
9. Maintenance, Local Taxes and Insurance - 4.5% of fixed investment
10. Product Values - all liquid fuel products are of equal value and are to be determined by this study
11. By-product values:

    (a) Ammonia at $120/short ton
    (b) Sulfur at $53/long ton

M. Miscellaneous

1. Use English units
2. Mass flow rates
    a. Barrels per stream day (BPSD)
    b. Short tons per day (ST/D)
    c. Standard cubic feet per day (SCFD)

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Discussion

The flow diagrams of the processing schemes producing JP-8 and JP-4 jet fuels are presented in the form of block flow diagrams noted as Figures 3 and 4. The purpose of the refinery is to process 100,000 BPSD of Occidental shale oil into acceptable aviation turbine fuels JP-4 and JP-8 and coproducts. The main upgrading features for the processing schemes are two stages of hydrotreating followed by hydrocracking which is the primary conversion unit for converting shale oil into valuable products.

Occidental Shale Oil

Two Green River formation shale oils were evaluated during Phase I and II. The primary feedstock, which is used in this evaluation, was Occidental shale oil derived from a modified in-situ retort. The other shale oil was Paraho which was derived from a direct heated above ground retort. Paraho shale oil is not a part of this evaluation. The inspection on the Occidental feedstock is shown in Table 102. The Occidental shale oil has a normal boiling range of 400-1000°F and a sulfur content of 0.64 wt-%. However, nitrogen and oxygen content and bromine numbers are much higher than conventional petroleum feedstocks. Typical with shale oils, the arsenic level for the Occidental feedstock is high with a concentration of 27.5 wt-ppm.

Process Units

The processing schemes for the production of JP-4 and JP-8 jet fuels are nearly identical except for the size of the units. A naphtha hydrotreating and Platforming® unit are added to the flow scheme when producing JP-8.

Hydrotreating (UOP RCD Unibon® and Hydrotreating)

The two stages of hydrotreating include both low and high pressure operations and are identical for both process schemes. The low pressure hydrotreater saturates the olefins and diolefins, and removes iron, arsenic and most of the sulfur compounds. A high concentration of nitrogen compounds is left in the product that must be removed in the high pressure hydrotreater. The low pressure hydrotreater product from the separator is not stripped or fractionated but pumped directly into the high pressure hydrotreater. The balance of the shale oil contaminants is removed with the nitrogen level taken down to less than 1000 wt-ppm. The combined low and high pressure hydrotreater yields are given in Table 103.

Hydrocracking (UOP HC Unibon)

The HC Unibon® proposed for this refinery is a parallel flow hydrocracker designed to process the hydrotreated Occidental shale oil into light ends, light and heavy naphthas and kerosine/diesel products. Two operational modes vary the end points on the kerosine to make the JP-4 or JP-8 jet fuels. Tables 104 and 105 present the yield estimates and product properties when operating for JP-4 and JP-8 jet fuels.

The hydrocracker is designed to operate under high pressure conditions and to crack the shale oil in the presence of hydrogen to yield more valuable, lower-boiling products. At the same time, the hydrocracking reactor almost completely converts the balance of the sulfur-, nitrogen-, and oxygen-containing compounds, thus yielding products that are mixtures of essentially only paraffins, naphthenes and aromatics. The kerosine products yielded from the HC Unibon are of good quality. Both kerosine cuts (JP-4 and JP-8) are nearly sulfur free, low in aromatics, have low freeze points and good smoke points. These hydrocracked kerosines make good jet fuel blending stock.

The light naphtha (C_5/C_6) is recovered from the naphtha splitter and taken to gasoline blending during JP-8 operation. When JP-4 is made, the light naphtha is blended with the heavy naphtha (C_7-300°F) and the 300-520°F kerosine to make the jet fuel. The light naphtha is a good gasoline blending stock with a clear Research Octane Number of 75. Naphtha splitter bottoms, heavy naphtha (C_7-300°F), is used directly for jet fuel blending (JP-4 case), or hydrotreated and reformed in a UOP Platforming unit to obtain a high octane gasoline blending component. Gasoline blending butanes are recovered from the C_4 minus light ends when making gasoline during JP-8 operation. The balance of the light ends are sent to fuel gas treating prior to being used as hyrogen plant feedstock. Capacities, utilities and investments can be found in Tables 109, 110 and 111.

Naphtha Hydrotreating (For JP-8 Case Only)

The C_7-300°F hydrocracked naphtha from the bottoms of the naphtha splitter is charged to the naphtha hydrotreater for complete desulfurization and other contaminant removal (oxygen and nitrogen) prior to reforming in a UOP Platforming unit. The treated naphtha is stripped and charged to the Platformer®. Table 106 presents the yield estimate for the naphtha hydrotreater which includes some of the product qualities. Tables 110 and 111 provide capacities, utilities and investments.

UOP Platforming (For JP-8 Case Only)

The estimated yields for the UOP Platforming unit are presented in Table 107. The Platformer is designed for processing 6,665 BPSD of naphtha in a conventional fixed bed, semi-regenerative type reformer, featuring low pressure operation with gas recontacting giving maximum liquid recovery and high hydrogen purity. The yields and some of the major properties are noted in Table 107. The unit is designed for 98 RON clear severity for the C_5+ Platformate®. The Platformer is designed for conventional catalyst regeneration with an estimated first cycle length of twelve months.

The liquid C_4- material from the Platformer debutanizer is sent to a depropanizer to recover butanes for gasoline blending. There is no attempt to recover propane, and the light end hydrocarbons are taken to the fuel gas system for feed to the hydrogen plant. Part of the hydrogen produced from the Platformer net separator gas is used in the naphtha hydrotreater. The balance, the larger portion, feeds the hydrogen plant. Tables 110 and 111 present the utilities and investment requirements for the UOP Platforming unit.

Hydrogen Plant

An estimated 250 million SCFD of pure hydrogen is required to satisfy all the hydrogen requirements for the JP-4 case refinery processing scheme which includes estimated mechanical and solution losses. Approximately 53.4 million SCFD of treated fuel gas is used as feedstock to the JP-4 case hydrogen plant. About 256 million SCFD of pure hydrogen is used in the JP-8 case processing scheme. The feedstock to this hydrogen plant is 49.5 million SCFD of treated fuel gas, 10.4 million SCFD of hydrogen rich Platforming gas and 1,515 BPSD of hydrocracked light naphtha. Table 108 presents the estimated yields for both JP-4 and JP-8 cases which includes types of feedstocks, hydrogen yields, purity and estimated process unit usage. Tables 109, 110 and 111 cover the hydrogen plant capacities, utilities and investments.

Treating Units

The net off-gases from the HC Unibon (hydrocracker) and the naphtha hydrotreater (JP-8 Case only) are amine treated for hydrogen sulfide removal prior to being used as feedstock for the hydrogen plant or fuel gas to the heaters. Most of the treated light gases are used as hydrogen plant feedstock with very little left over for heating.

The gaseous hydrogen sulfide from the amine regenerator and sour water stripper are combined and sent to the Claus sulfur plant where elemental sulfur is recovered.

Sour water from the low and high pressure hydrotreater, from the hydrocracker and from the naphtha hydrotreater are treated in the sour water stripper. Sour water contains hydrogen sulfide, ammonia and other washed out contaminants. Stripped water is recovered and the released gases, hydrogen sulfide and ammonia are separated with hydrogen sulfide, processed in the sulfur plant and the ammonia recovered in a separation plant.

Recovered Products

The overall material balances for a refinery producing aviation turbine fuel from 100,000 BPSD of raw shale oils are given in Tables 112 and 113.

Aviation Turbine Fuels -- The refinery is designed to produce two types of aviation turbine fuels as the major required products. These are:

A. JP-4 Jet Fuel -- This military jet fuel is of the gasoline type with a distillation range from C_5-518°F with gravity limitations between 45-57 °API. Table 114 compares the product qualities with the Military Specifications for JP-4 and the jet fuel does meet military specifications. Tabulated below is the composition of the JP-4 jet fuel.

Component	BPSD	ST/D	Sp. Gr.	°API
Lt. Naph. (C_5/C_6)	13,399	1,556.9	0.6637	81.7
C_7-300°F Naph.	7,239	965.0	0.7614	54.3
300-520°F Kerosine	73,780	10,354.8	0.8016	45.0
Total	94,418	12,876.7	0.7790	50.2

B. JP-8 Jet Fuel -- This military jet fuel is of the kerosine type similar to ASTM Jet A-1 fuel with a maximum distillate end point of 572°F with gravity limitations between 37-51 °API. Table 115 compares the estimated product with the JP-8 military

specification and the jet fuel product does meet or exceed these specifications. The composition of the JP-8 jet fuel is shown below:

Component	BPSD	ST/D	Sp. Gr.	°API
300-550°F Kerosine	82,406	11,609.2	0.8047	44.3

Motor Gasoline -- Motor gasolines are produced only when JP-8 jet fuels are made. The motor gasoline blended in this study is special grade unleaded which meets Military Specifications and Federal Specification VV-G-1690B dated July 1, 1978. Table 116 compares the study blend with the required specifications and the blend equals or exceeds these specifications. The composition of the gasoline blend is as follows:

Component	BPSD	ST/D	Sp. Gr.	RVP	RON Clear	MON Clear	R+M/2
Platformate	5,414	773.3	0.8158	3.9	98.0	87.0	
Lt. Naphtha	1,887	220.0	0.6660	11.0	75.0	73.0	
Mixed Butanes	814	81.8	0.5738	65.0	98.8	93.8	
Total	8,115	1,075.1	0.7567	12.6	92.6	84.4	88.5

It should be noted that the octanes exceed the specifications and the vapor pressure is below the maximum of 13.5 psi. The composition of the blend is toward a light gasoline with 1/3 of the blend composed of light gasoline and butanes, but the blend is still acceptable by distillation specifications.

Refinery Fuel Gas -- As the material balances indicate, very little of the fuel gas is used for heating in each of the processing schemes. Table 112, the material balance for the JP-4 case, indicates that very little gas is left for fuels and that most of the treated fuel gas was used in the hydrogen plant. Refering to Table 113, the material balance for JP-8, all of the fuel gas was sent to the hydrogen plant.

Refinery Fuel Oil -- As noted above, all the heating will be done with fuel oil. Fuel oil is drawn from the refinery processing scheme after the high pressure hydrotreater. High pressure separator liquid, drawn for fuel oil, is flashed with the light hydrocarbon material going to fuel gas treating and the stabilized liquid used in the fuel oil system. For the JP-4 case, 13,508 BPSD of fuel oil is drawn from the high pressure separator which represents about 13-1/2% of the shale oil feed liquid volume. For the JP-8 case about 7179 BPSD of fuel oil is drawn plus 9075 BPSD of light naphtha which can be used as fuels in the hydrogen plant.

Overall Refinery Flow Description

The processing schemes for making JP-4 and JP-8 jet fuels from shale oil are essentially the same. The major process units common to both processing schemes are as follows: 1) Shale oil desalting, 2) Low and high pressure hydrotreating, 3) Hydrocracking of the hydrotreated shale oil, 4) Fractionation of the hydrocracked products and stabilization of the light ends and naphthas, 5) Hydrogen production, and 6) Treating (amine treating, sulfur plant and sour water stripping). Naphtha hydrotreating and Platforming are used only in the JP-8 jet fuel case.

Low Pressure Hydrotreating (RCD Unibon)

The desalted shale oil charge stock is combined with amine scrubbed recycle hydrogen gas plus make-up hydrogen gas (from the hydrogen plant), heated to the desired reaction temperature by means of exchange and direct fired heaters and charged to the reactors. The reactor effluent, after heat exchange with the incoming feed, is water washed, condensed, cooled and sent to the product separator. The hydrogen rich off-gas from the separator is amine treated and recycled to the reactor section. Ammonia-rich water is removed from the separator and sent to the sour water stripper. Liquid hydrocarbon product from the separator is pumped directly to the high pressure hydrotreater for further processing.

The low pressure hydrotreater saturates the olefins and diolefins by contacting the feedstock with high purity hydrogen on a catalyst bed. Iron and arsenic are removed along with most of the sulfur compounds. Essentially no arsenic and only small amounts of unsaturates remain; however, a high concentration of nitrogen compounds is still present in the first stage product and must be removed in the high pressure hydrotreating step.

High Pressure Hydrotreating

The first-stage low pressure separator liquid is combined with the recycle hydrogen rich gas plus make-up hydrogen (from the hydrogen plant), heated to the desired reactor temperature by means of exchange and direct fired heaters, and charged to the reactors. The reactor effluent is heat exchanged with the reactor feed, water washed, condensed, cooled and sent to the product separator. The hydrogen rich gas from the separator is recycled to the reactor, and the ammonia rich water is drawn from the separator and treated in the sour water stripper. Liquid hydrocarbons from the separator are pressured directly to the hydrocracking unit (HC Unibon) for further processing.

The high pressure hydrotreating step completes the denitrification of the shale oil to less than 1000 wt-ppm. Hydrocarbon liquid to be used as fuel oil for process heaters is drawn from the high pressure hydrotreater separator. This material will be stripped, with the gas going to fuel gas treating and the liquid used as fuel oil. The balance of the separator liquid will be sent on to the hydrocracker.

Hydrocracking (HC Unibon)

Hydrocracking is a highly versatile process for the conversion of a variety of petroleum/hydrocarbon fractions to yield more valuable, lower-boiling products. Along with the molecular weight reduction accomplished in the hydrocracking reactions, there is almost complete conversion of sulfur-, nitrogen-, and oxygen-containing compounds, thus yielding

products that are mixtures of essentially only paraffins, naphthenes and aromatics. The refinery study requires a parallel flow hydrocracker to process the hydrotreated shale oil charged from the high pressure hydro-treater separator.

The hydrotreated shale oil charge stock is combined with the recycle hydrogen rich gas plus hydrogen gas make-up, heated to the desired reactor temperature by means of heat exchange and a direct fired heater, and charged to the reactor section. The reactor section effluent is water washed, cooled and sent to the product separator. The hydrogen rich off-gas from the separator is returned to the reactor section as recycle gas and hydrogen quench.

Ammonia is removed from the product separator with the wash water. The hydrocarbon liquid from the separator is flashed to the low pressure flash drum where much of the H_2S and light gases are removed to the fuel gas treating system. The low pressure flash hydrocarbon liquid is charged to a fired heater and sent to the product fractionator where the IBP-300°F material is taken overhead, the 300-520°F kerosine is stripped and blended with the C_5+ naphtha product from the overhead material into JP-4 jet fuel. For the JP-8 jet fuel case, the 300-550°F kerosine is stripped for flash point and sent to product storage. If diesel products are required with the jet fuels, a lower side cut draw is made on the product fraction-ator and a side cut stripper must also be provided to make flash point. The product fractionator bottoms material is combined with recycle hydro-gen gas and make-up hydrogen and charged to the reactor section.

Light Ends Fractionation

The product fractionator overhead material is cooled and condensed then collected in the overhead receiver. The overhead receiver gas is compressed and recontacted with the overhead naphtha liquid in the recon-tact drum. Net gas is removed to fuel gas treating, and the naphtha is sent to the naphtha splitter, where C_6 minus material is taken overhead and the C_7-300°F heavy naphtha is taken as bottoms product for blending into the JP-4 jet fuel or to a UOP Platforming unit for making high

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octane gasoline, when the plant is making JP-8 jet fuel. The naphtha splitter overhead (C_6-) material is charged to a debutanizer to remove C_4- material from the light naphtha (C_5/C_6). The debutanized light naphtha is blended into the jet fuel for the JP-4 jet fuel case to make the gasoline type jet fuel, or blended into the gasoline pool with high octane Platformate produced during the JP-8 jet fuel production. If required, part of the light naphtha can be used in the hydrogen plant as fuels/feedstock. A small depropanizer is used to separate the C_3 and lighter material from the butanes in the C_4- overhead liquid from the debutanizer. The bottoms butanes are used to make vapor pressure in the gasoline during the JP-8 operation.

Naphtha Hydrotreating (JP-8 Case Only)

In the hydrotreating process unit the $C_7-300°F$ hydrocracked naphtha obtained from the naphtha splitter is combined with recycle hydrogen rich gas plus make-up hydrogen gas (from the Platformer), heated to the desired reaction temperature by means of exchange and direct fired heater and charged to the reactor. The effluent from the reactor, after heat exchange, with the incoming feed, is condensed, cooled and sent to the product separator. The hydrogen rich off-gas from the separator is recycled to the reactor section while net separator gas is sent to fuel gas.

The liquid product from the separator is sent to a stripper column for hydrogen sulfide and light ends removal. The C_6 plus stripper bottoms product is charged directly to the Platformer.

UOP Platforming (JP-8 Case Only)

In the UOP semi-regenerative Platforming unit the hydrotreated naphtha from the naphtha hydrotreater is combined with the recycle hydrogen rich gas stream, heated by means of exchange and a fired heater, then charged successively through the reforming reactors. Fired interheaters are provided between the reactors in order to achieve the desired reactor inlet temperatures.

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The effluent from the last reactor, after heat exchange with the incoming feed, is condensed, cooled and sent to the product separator. The hydrogen rich gas stream from the separator is recycled to the reactor section while the off-gas produced is used as make-up hydrogen for the naphtha hydrotreater and as feedstock to the hydrogen plant. The separator hydrocarbon liquid after recontacting with the compressed separator gas, is separated and charged to the stabilizer (debutanizing for motor fuel operation) where the light hydrocarbons are taken overhead to hydrocracker fractionation and the C_5^+ Platformate is sent to gasoline blending.

Amine Treating and Sulfur Recovery

Amine treating of refinery gases removes hydrogen sulfide and other acid gas constituents avoiding atmospheric pollution by SO_2 from combustion of these gases. The amine treating process is based on an absorption-regeneration cycle using aqueous solutions of an alkanolamine which reacts with acid gases. Hydrogen sulfide containing feed is contacted countercurrently with amine solution in an absorption or extraction column. Regenerated amine solution is fed into the top of the column and the rich solution taken from the bottoms through exchange to regeneration, where the acid gases are stripped with steam and recovered as hydrogen sulfide with the regenerated solution recycled back to the absorber. The hydrogen sulfide from the amine unit and sour water stripper is charged to a Claus sulfur recovery unit for conversion to elemental sulfur.

Sour Water Treating

In the hydrocracking and hydrotreating processing schemes, water is injected to prevent build-up of salts on the surface of heat exchange tubes. Wash water used to dissolve ammonia and some hydrogen sulfide cannot be disposed of in the conventional refinery sewer system because of its chemical content. The waste water is charged to a stripper, where hydrogen sulfide and ammonia are steam stripped from the water. The clean water is returned to the refinery and the gas is sent to the amine treater for further treatment.

Hydrogen Plant

The basic steps for the steam-hydrocarbon reforming process in the production of hydrogen are desulfurization, reforming, two-stage CO conversion, CO_2 removal, methanation and compression to the required pressure levels.

A variety of feedstocks can be used, such as natural gas, methane, ethane, propane and butanes. Liquid feeds such as light naphthas are sometimes used. The feed should contain less than 0.5% olefins to prevent carbon formation. The hydrocarbons are charged to the charcoal absorber for sulfur removal then sent to the reformer furnace with steam. The gas is converted largely to hydrogen, CO and CO_2. The gases are cooled by the addition of steam or condensate and all passed over catalyst in the CO converter. About 90 to 95% of the CO is converted to CO_2. The hot gases leaving the converter are cooled and then scrubbed with an amine solution to remove most of the CO_2. The gases are heated up after leaving the CO_2 absorber by exchange with gas coming from the CO converter and pass over catalyst in the methanator. Here all the carbon oxides are converted to methane by reaction with hydrogen. The product hydrogen gas from the methanator, saturated with water vapor, is cooled to 100°F.

Capital Costs

The economic evaluation was performed using the basis established in the Basis and Scope of Study section with one exception: that the capital cost basis used was for the 1st quarter, 1981. In Table 111, the estimated erected cost (EEC) for individual process units is shown. The costs for the JP-4 and JP-8 cases are similar with one exception: a naphtha hydrotreater and Platformer unit combination is included only in the maximum JP-8 case. As shown in Table 117, Capital Investment Summary, the total erected cost for the maximum JP-4 refinery is $526 Million and the maximum JP-8 refinery is $556 million. In addition, the offsites allowance for both schemes is $150 million. This allowance assumes that the shale oil processing facilities would be constructed in conjunction with

an existing oil refinery. The adjustment of this allowance upward to be consistent with a grassroots shale oil refinery will be investigated during Phases III and IV of this project. The allowance for fully paid royalties is added to the plant investment to obtain the total depreciable investment, which is $695 million for JP-4 case and $726 million for the JP-8 case.

Operating Costs

The overall refinery operating costs are presented in Table 118 for the JP-4 case and in Table 119 for the JP-8 case. Operating costs are divided into direct and indirect operating costs which are as follows:

I. Direct Operating Cost

1. **Refinery Labor** - Refinery labor includes all the personnel for the process units hired at $12.00 per hour for operators and $10.50 per hour for helpers, and a 35% fringe is added to these base rates. A 25% allowance for supervision is added plus 100% for labor overhead.

2. **Maintenance Allowance** - The maintenance allowance covers normal operating maintenance and turnaround contract maintenance for all refinery equipment including process units, offsites and depreciable assets. An amount equal to 3% of erected plant investment is allocated for maintenance.

3. **Utilities** - Refinery fuel is generated internally for both JP-4 and JP-8 cases. Power is purchased at $0.035 per kWh and cooling water is priced at $0.03 per 1,000 gallons. Boiler feed water is estimated at $0.30 per 1,000 lb. Steam (600 psig) is credited at $5.00 per 1,000 lb. All utilities are consumed at normal average operating rates.

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4. Catalyst, Solvents and Chemicals - Catalyst consumption is based on the expected catalyst life for normal operating conditions. Similarly, solvents and chemicals are based on normal average operating usage.

II. Indirect Operating Cost

1. Local Taxes - An allowance of 1% of erected plant investment is allocated toward local taxes.

2. Insurance - An allowance of 1/2% of erected plant investment is allocated toward insurance.

Economic Analysis

Incorporating the data generated during the pilot plant operations, material balances for a shale oil refinery processing Occidental shale oil were prepared and liquid fuel production costs were calculated. The results that follow are considered preliminary; the preparation of more accurate estimates is a part of future work under this contract.

Overall Material Balances

Overall material balances have been prepared for two cases: maximum JP-4 production and maximum JP-8 production. These maxima were established by selecting appropriate cut points for product fractionation operation. In the JP-8 case, based on the results of the pilot plant operations, an end point less than the military specification maximum was required in order to meet the freeze point specification. In the JP-4 case, maximum military specification end point was possible without compromising any other specification. The resulting overall material balances are presented in Tables 112 and 113 for maximum JP-4 and maximum JP-8 cases, respectively. These balances are based on 100 wt-% shale oil feed. Hydrogen is generated internally as are all process fuel and heat requirements. Only process water, cooling water and electric power are required to be supplied externally.

Hydrogen requirements for both cases are similar with a 250 MMSCFD hydrogen plant for maximum JP-4 case and a 256 MMSCFD hydrogen plant for maximum JP-8 case. (See Table 108.) Jet fuel yields of 80.25 wt-% (94.42 vol-%) are shown in the JP-4 case and 72.35 wt-% (82.40 vol-%) in the JP-8 case. The JP-4 case produces no other liquid fuel product, whereas there is 6.70 wt-% (8.12 vol-%) of gasoline produced in the JP-8 case. The total liquid fuel yields from the two refinery cases then are 80.25 wt-% (94.42 vol-%) and 79.05 wt-% (90.52 vol-%) for maximum JP-4 and maximum JP-8, respectively.

Production Costs

The total costs of production are tabulated in the Proforma Financial Statements for the JP-4 and JP-8 cases in Tables 120 and 121, respectively. These statements assume that all liquid products have equal value. Including feedstock cost, operating costs, capital charges, taxes and a 15% DCF rate, the total cost of liquid fuel production from the maximum JP-4 refinery is $40.57 per barrel of shale oil feed and from the maximum JP-8 refinery is $41.10 per barrel of feed. The difference in cost is almost entirely due to the higher capital cost of the JP-8 refinery.

The production costs can be stated on "per barrel of total liquid fuel" basis as noted in the Proforma Financial Statements, pages 3-1 of Tables 120and 121. This is calculated by dividing production costs by the volume fraction yield of liquid fuel. This calculation results in total liquid fuel costs of $42.97 and $45.40 per barrel for the maximum JP-4 and maximum JP-8 cases, respectively.

TABLE 102. SHALE OIL INSPECTION

Occidental Shale Oil

Feed Source	Desalted Shale Oil
Gravity, °API	22.9
Sulfur, wt-%	0.64
Nitrogen, wt-%	1.51
Pour Point, °F	+65
BS and W, vol-%	0.2
Conradson Carbon, wt-%	1.36
Carbon, wt-%	79.3
Hydrogen, wt-%	12.2
Oxygen, wt-%	0.65
Metals, wt-ppm	
Arsenic	27.5
Iron	23.0
Nickel	10.0
Vanadium	5.0
Ash, wt-%	0.014
Bromine Number	23.5
ASTM Distillation, °F	
IBP	376
50%	712
EP	953
% Over	87

TABLE 103. OVERALL REACTOR YIELDS

Low and High Pressure Hydrotreating

Feed	Maximum JP-4 and JP-8	
	Wt-%	(SCFB) LV-%
Shale Oil	100.00	100.00
Hydrogen (Chemical)	2.36	(1425)
Total	102.36	

Products

Water	0.73	
Ammonia	1.85	
Hydrogen Sulfide	0.68	
Propane and Lighter	0.76	
Butanes	0.32	0.51
Pentanes	0.53	0.78
Hexane +	97.49	103.13
Total	102.36	104.42

Estimated Properties of C_6+

°API	31.8
Sulfur, wt-%	0.003
Nitrogen, wt-%	0.09
C_7 Insolubles, wt-%	< 0.05
Bromine Number	1.1
ASTM Dist., °F	
IBP	245
50%	646
EP	1034

TABLE 104. REACTOR YIELDS

HC Unibon

Feed	Maximum JP-4	
	Wt-%	(SCFB) LV-%
HT Shale Oil*	100.00	100.00
Hydrogen (Chemical)	1.57	(885)
Total	101.57	

Products

	Wt-%	LV-%
Ammonia	0.11	
Propane and Lights	3.53	
Butanes	4.85	
Light Naphtha (C_5/C_6)	11.25	14.50
Heavy Naphtha (C_7-300°F)	6.98	7.83
Kerosine (300-520°F)	74.85	79.84
Total	101.57	102.17

Estimated Properties

	Charge	C_5/C_6	Naphtha	Kerosine
Dist. Range, °F			C_7-300	300-520
°API	31.8	80.5	54.3	45.0
Sulfur, wt-ppm	30	3	3	3
Nitrogen, wt-ppm	900	< 1	< 1	3
ASTM Dist., °F				
IBP	245	90	200	300
50%	646	130	240	404
EP	1034	180	285	520
RVP, psi	-	12.9	-	-
P/N/A, vol-%	-	83/15/2	41/49/10	-
RON Clear	-	75	59	-
RON + 3 cc TEL/USG	-	93	78	-
Flash Point, °F	-	-	-	100
Aromatics, vol-%	-	-	-	13.8
Freeze Point, °F	-	-	-	-58
Naphthalenes, wt-%	-	-	-	0.4
Smoke Point, mm	-	-	-	25

* Total hydrotreater, high pressure separator liquid.

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TABLE 105. REACTOR YIELDS

HC Unibon

Feed	Maximum JP-8	
	Wt-%	(SCFB) LV-%
HT Shale Oil*	100.00	100.00
Hydrogen (Chemical)	1.55	(874)
Total	101.55	

Products

	Wt-%	LV-%
Ammonia	0.11	
Propane and Lighter	3.13	
Butanes	4.25	
Light Naphtha (C_5/C_6)	9.78	12.60
Heavy Naphtha (C_7-300°F)	5.99	6.73
Kerosine (300-550°F)	78.29	83.20
Total	101.55	102.53

Estimated Properties

	Charge	C_5/C_6	Naphtha	Kerosine
Dist. Range, °F			C_7-300	300-550
°API	31.8	81.7	54.3	44.3
Sulfur, wt-ppm	30	3	3	3
Nitrogen, wt-ppm	900	< 1	< 1	3
ASTM Dist., °F				
IBP	245	90	200	300
50%	646	130	240	419
EP	1034	180	285	550
RVP, psi	-	12.9	-	-
P/N/A, vol-%	-	83/15/2	41/49/10	-
RON Clear	-	75	59	-
RON + 3 cc TEL/USG	-	93	78	-
Flash Point, °F	-	-	-	100
Aromatics, vol-%	-	-	-	19
Freeze Point, °F	-	-	-	-58
Naphthalenes, wt-%	-	-	-	1.5
Smoke Point, mm	-	-	-	23

* Total hydrotreater, high pressure separator liquid.

TABLE 106. REACTOR YIELDS

Naphtha Hydrotreating Unit

	Maximum JP-8	
Feed	**Wt-%**	**(SCFB) LV-%**
Hydrocracked Naphtha (C_7-300°F)	100.00	100.00
Hydrogen (Chemical)	0.03	(16)
Total	100.03	

Products

Hydrogen Sulfide	0.02	
Propane and Lighter	0.02	
Butanes	0.01	
Pentanes	0.02	
Hexane +	99.96	100.01
Total	100.03	100.01

Properties of Hexane +

°API	54.4
Sulfur, wt-%	0.00005
Hydrocarbon Type, vol-%	
P	41
N	49
A	10

TABLE 107. REACTOR YIELDS

UOP Platforming Unit

	Maximum JP-8	
Feed	**Wt-%**	**(SCFB)** **LV-%**
Hydrotreated Naphtha (C_7-300°F)	100.00	100.00

Products

Hydrogen	2.54	(1273)
Propane and Lighter	6.26	-
Butanes	3.53	4.67
C_5+ Platformate	87.67	81.85
Total	100.00	

Properties of C_5+ Platformate

°API	41.9
RON Clear	98.0
RON + 3 cc TEL	103.7
MON Clear	87.0
MON + 3 cc TEL	91.6
RVP, psi	3.9

TABLE 108. ESTIMATED YIELDS

Hydrogen Plant

Feed to H_2 Plant	JP-4 Case					JP-8 Case				
	BPSD	Sp.Gr.	MW	MMSCFD	Wt-%	BPSD	Sp.Gr.	MW	MMSCFD	Wt-%
Treated Fuel Gas			21.9	53.4	100.0			18.9	49.5	81.0
Platformer Gas								8.4	10.4	7.5
Light Naphtha						1515	0.6638	-	-	11.5
Total Feed					100.0					100.0
Products										
Hydrogen				250.0	43.0				256.0	44.6
CH_4				7.7	10.6				7.9	11.0
Total H_2 (97% Purity)				257.7	53.6				263.9	55.6
Process Loss					46.4					44.4
Total Products					100.0					100.0

Estimated Process Units H_2 Requirements

Unit	H_2 MMSCFD	H_2 MMSCFD
Low and High Pressure Hydrotreater	152.7	152.7
Hydrocracker (HC Unibon)	95.1	100.9
Naphtha Hydrotreater	-	0.2
Mechanical Losses	2.2	2.2
Total Hydrogen Required	250.0	256.0

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TABLE 109. UTILITY CONSUMPTION

Maximum JP-4

	Power, kWh	Steam, M lb/hr			BFW and Process Water, M lb/hr	Cooling Water, gpm	Fuel Fired, MM btu/hr
		HP	MP	LP			
LP Hydrotreating (RCD Unibon)	8,822	-	-	-	-	37	62
HP Hydrotreating	19,443	-	-	-	-	112	177
HC Unibon – Hydrocracking	22,467	-31.1	-1.1	-	-22.4	1482	724
Hydrogen Plant (Steam Reforming)	8,308	+172.5	-	-	-571.2	500	1965
Amine Treating	147	-	-	-27.7	-	122	-
Sulfur Plant	54	-0.8	+11.8	+27.0	-38.2	-	13
Sour Water Stripper	353	-	-81.9	-	-	-	-
Fractionation	235	-	-34.2	-5.9	+40.1	662	-
Feed Preparation	150	-	-	-	-	-	-
TOTAL	59,979	+140.6	-105.4	-6.6	-591.7	2915	2941

NOTES:

1. Negative sign indicates consumption
2. Positive sign indicates production
3. Maximum air cooling
4. Fuel fired heating based on net heating value.

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TABLE 110. UTILITY CONSUMPTION

Maximum JP-8

| | Power, kWh | Steam, M lb/hr | | | BFW and Process Water, M lb/hr | Cooling Water, gpm | Fuel Fired, MM btu/hr |
		HP	MP	LP			
LP Hydrotreating (RCD Unibon)	8,822	-	-	-	-	37	62
HP Hydrotreating	19,443	-	-	-	-	112	177
HC Unibon - Hydrocracking	23,967	-35.9	+2.1	-	-24.7	1514	781
Naphtha Hydrotreating	230	-	-	-	2.9	122	20
Platforming	1,200	+24.9	-	-	26.2	504	97
Hydrogen Plant (Steam Reforming)	8,180	+190.3	-	-	-605.4	457	2014
Amine Treating	147	-	-	-27.7	-	122	-
Sulfur Plant	54	-0.8	+11.8	+27.0	-38.2	-	13
Sour Water Stripper	359	-	-83.5	-	-	-	-
Fractionation	251	-	-36.6	-6.3	+42.9	708	-
Feed Preparation	150	-	-	-	-	-	-
TOTAL	62,803	+178.5	-106.2	-7.0	-654.6	3576	3191

NOTES:

1. Negative sign indicates consumption
2. Positive sign indicates production
3. Maximum air cooling
4. Fuel fired heating based on net heating value.

TABLE 111. PROCESS UNITS CAPACITIES AND CAPITAL INVESTMENTS

	Max. JP-8		Max. JP-4	
	(MMSCFD) BPSD	MM Dollars	(MMSCFD) BPSD	MM Dollars
LP Hydrotreating (RCD Unibon)	100,000	44	100,000	44
HP Hydrotreating	102,924	125	102,924	125
Hydrocracking (HC Unibon)	102,500	198	95,140	188
Naphtha Hydrotreating	6,665	4		-
Platforming	6,665	13	-	-
Hydrogen Plant	(256)	154	(250)	151
Amine Treating		7		7
Sulfur Plant		7		7
Sour Water Stripper		2		2
Feed Preparation		2		2
Total Process Investment		556		526

Note: Capital Investments as of 1st quarter 1981

TABLE 112. OVERALL MATERIAL BALANCE

JP-4 Jet Fuel

Feed	Maximum JP-4		
	Wt-%	Vol-%	BPSD
Shale Oil	100.00	100.00	100,000
Products			
Gasoline	-	-	-
Jet Fuel JP-4	80.25	94.42	94,418
Subtotal	80.25	94.42	94,418
By-Products			
Sulfur	0.63	-	-
Ammonia	1.96	-	-
Water	0.73	-	-
Subtotal	3.32	-	-
Streams Utilized as Fuel			
Fuel Gas	0.28	-	-
Fuel Oil	12.74	13.51	13,508
Subtotal	13.02	13.51	13,508
Material Lost in Production	3.41		
GRAND TOTAL	100.00		

TABLE 113. OVERALL MATERIAL BALANCE

JP-8 Jet Fuel

Feed	Maximum JP-8		
	Wt-%	Vol-%	BPSD
Shale Oil	100.00	100.00	100,000
Products			
Gasoline	6.70	8.12	8,115
Jet Fuel JP-8	72.35	82.40	82,406
Subtotal	79.05	90.52	90,521
By-Products			
Sulfur	0.63	-	-
Ammonia	1.96	-	-
Water	0.73	-	-
Subtotal	3.32	-	-
Streams Utilized as Fuel			
Light Naphtha	6.57	9.07	9,075
Fuel Oil	6.70	7.18	7,179
Subtotal	13.27	16.25	16,254
Material Lost in Production	4.36		
GRAND TOTAL	100.00		

TABLE 114. PRODUCT QUALITIES

JP-4 Jet Fuel

	JP-4 Military Specs.	Max. JP-4
Gravity, °API	45-57	50.2
Aromatics, vol-%	25 (Max.)	11.9
Flash Point, °F	-	-
Freeze Point, °F	-72 (Max.)	< -72
Smoke Point, mm	20	31
Sulfur, wt-%	0.40	0.0003
Distillation, °F		
IBP	Report	107
10%	Report	175
20%	293 (Max.)	250
50%	374 (Max.)	372
90%	473 (Max.)	473
EP	518 (Max.)	518

TABLE 115. PRODUCT QUALITIES

JP-8 Jet Fuel

	JP-8 Military Specs	Max. JP-8
Gravity, °API	37-51	44.3
Aromatics, vol-%	25 (Max)	19
Flash Point, °F	100 (Min)	100
Freeze Point, °F	-58 (Max)	-58
Smoke Point, mm	20 (Min.)	23
Sulfur, wt-%	0.40	0.0003
Naphthelenes, vol-%	3.0 (Max)	1.5
Distillation, °F		
IBP	Report	300
10%	401 (Max)	337
20%	Report	361
50%	Report	419
90%	Report	512
EP	572 (Max.)	550

TABLE 116. GASOLINE BLEND AND QUALITIES

Special Grade -- Unleaded

Gasoline Blend	Maximum JP-8	
	Vol-%	BPSD
Platformate	66.7	5,414
Light Gasoline (C_5/C_6)	23.3	1,887
Mixed Butanes	10.0	814
TOTAL	100.00	8,115

Properties of Gasoline	Military Specs.	Max. JP-8 Gasoline
°API	-	55.5
RON	92.0 (Min.)	92.6
MON	82.0 (Min.)	84.4
R + M/2	87.0 (Min.)	88.5
RVP, psig	13.5 (Max.)	12.6
Sulfur, wt-%	0.10 (Max.)	0.00005

ASTM Dist., °F

IBP	-	104
10%	131 (Max.)	122
30%	-	150
50%	171 (Min.) - 235 (Max.)	178
70%	-	233
90%	365 (Max.)	283
EP	437 (Max.)	336

TABLE 117. CAPITAL INVESTMENT SUMMARY

(Million of Dollars)

100,000 BPSD Refinery	JP-8	JP-4
Process Units Erected Cost	556	526
Allowance for Offsites	150	150
Allowance for Fully Paid Royalties and Know-How Fees	20	19
Total Depreciable Investment	726	695
Initial Catalyst Inventory	9	9
Working Capital Allowance	67	68
Total Capital Investment	802	772

TABLE 118. JP-4 CASE -- ESTIMATED OPERATING COST

Basis: 330 Operating Days

Direct Operating Cost

		MM$/Yr.	MM$/Yr.
A.	**Refinery Labor**		
	Direct Labor		
	70 operators at $38,316/Person/Year	2.680	
	40 helpers at $33,527/Person/Year	1.000	
	Total Direct Labor	3.680	
	Supervision		
	25% of Direct Labor	0.925	
	Operating Labor	4.605	
	Overhead		
	100% of Operating Labor	4.605	
	Total Labor	9.210	9.21
B.	**Maintenance**		
	Maintenance Allowance at 3% of Plant Investment		20.28
C.	**Utilities**		
	Power: 59,979 kWh at $0.035/kWh	16.63	
	Boiler Feed Water: 591.7 M lb/hr at $0.30/M lb	1.40	
	Cooling Water: 2915 gpm at $0.03/1,000 gal	0.04	
	Steam (credit): 28.6 M lb/hr at $5.00/M lb	(1.13)	
	Total Utilities	16.94	16.94
D.	**Catalyst Replacement, Solvents and Chemicals**		
	Low Pressure Hydrotreating (RCD Unibon)	3.73	
	High Pressure Hydrotreating	1.41	
	Hydrocracking (HC Unibon)	2.28	
	Hydrogen Plant	1.50	
	Total Catalysts, Solvents and Chemicals	8.92	8.92
	Total Direct Operating Cost		55.35

TABLE 118. JP-4 CASE -- ESTIMATED OPERATING COST (Continued)

	MM$/Yr.
Indirect Operating Cost	
Local Taxes	
Local Taxes Allowance at 1% of Plant Investment	6.76
Insurance	
Insurance Allowance at 1/2% of Plant Investment	3.38
Total Indirect Operating Cost	10.14
Total Operating Cost	65.49

TABLE 119. JP-8 CASE --ESTIMATED OPERATING COST

Basis: 330 Operating Days

Direct Operating Cost

A. Refinery Labor

		MM$/Yr.	MM$/Yr.

Direct Labor
80 operators at $38,316/Person/Year   3.062
35 helpers at $33,527/Person/Year   1.173

Total Direct Labor   4.235

Supervision

25% of Direct Labor   1.065

Operating Labor   5.295

Overhead

100% of Operating Labor   5.295

Total Labor   10.590   10.59

B. Maintenance

Maintenance Allowance at 3%
of Plant Investment   21.18

C. Utilities

Power: 62,803 kWh at $0.035/kWh   17.409
Boiler Feed Water: 645.6 M lb/hr at $0.30/M lb   1.555
Cooling Water: 3576 gpm at $0.03/1,000 gal   0.051
Steam (credit): 65.3 M lb/hr at $5.00/M lb   (2.586)

Total Utilities   16.429   16.43

TABLE 119. JP-8 CASE -- ESTIMATED OPERATING COST (Continued)

	MM$/Yr.	MM$/Yr.
D. Catalyst Replacement, Solvents and Chemicals		
Low Pressure Hydrotreating (RCD Unibon)	3.73	
High Pressure Hydrotreating	1.41	
Hydrocracking (HC Unibon)	2.34	
Naphtha Hydrotreating	0.01	
UOP Platforming	0.06	
Hydrogen Plant	1.55	
Total Catalysts, Solvents and Chemicals	9.10	9.10
Total Direct Operating Cost		57.30

Indirect Operating Cost

Local Taxes

Local Taxes Allowance at 1% of Plant Investment		7.06

Insurance

Insurance Allowance at 0.5% of Plant Investment		3.53
Total Indirect Operating Cost		10.59
Total Operating Cost		67.89

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TABLE 120
USAF SHALE OIL STUDY
PROFORMA FINANCIAL STATEMENT

ECONOMIC EVALUATION MODEL

RUN NUMBER 1

JP4:15%ROI

CASE JP4
PAGE 1 - 1

ALL $ VALUES IN MILLIONS

YEAR	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
PERCENT CAPACITY	50.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
ANNUAL SALES	676.20	1352.40	1352.40	1352.40	1352.40	1352.40	1352.40	1352.40	1352.40	1352.40
RAW MATERIAL COST	495.00	990.00	990.00	990.00	990.00	990.00	990.00	990.00	990.00	990.00
GROSS MARGIN	181.20	362.40	362.40	362.40	362.40	362.40	362.40	362.40	362.40	362.40
OPERATING LABOR	9.21	9.21	9.21	9.21	9.21	9.21	9.21	9.21	9.21	9.21
UTILITY COST	8.47	16.94	16.94	16.94	16.94	16.94	16.94	16.94	16.94	16.94
MAINTENANCE COST	20.28	20.28	20.28	20.28	20.28	20.28	20.28	20.28	20.28	20.28
CAT+CHEM COST	4.46	8.92	8.92	8.92	8.92	8.92	8.92	8.92	8.92	8.92
DIRECT MFG EXPENSE	42.42	55.35	55.35	55.35	55.35	55.35	55.35	55.35	55.35	55.35
INSURANCE	3.38	3.38	3.38	3.38	3.38	3.38	3.38	3.38	3.38	3.38
PROPERTY TAX	6.76	6.76	6.76	6.76	6.76	6.76	6.76	6.76	6.76	6.76
INDIRECT MFG EXPENSE	10.14	10.14	10.14	10.14	10.14	10.14	10.14	10.14	10.14	10.14
DEPRECIATION	99.29	91.65	84.01	76.37	68.74	61.10	53.46	45.82	38.19	30.55
INTEREST ON W. CAPITAL	9.24	9.24	9.24	9.24	9.24	9.24	9.24	9.24	9.24	9.24
TAXABLE INCOME	20.11	196.02	203.66	211.29	218.93	226.57	234.21	241.84	249.48	257.12
TAX LOSS DEDUCTION	20.11									
ADJ. TAXABLE INCOME	20.11	196.02	203.66	211.29	218.93	226.57	234.21	241.84	249.48	257.12
CURRENT TAX	10.06	98.01	101.83	105.65	109.47	113.28	117.10	120.92	124.74	128.56
TAX CREDITS	9.05	69.91	0.	0.	0.	0.	0.	0.	0.	0.
TAXES PAYABLE	1.00	28.10	101.83	105.65	109.47	113.28	117.10	120.92	124.74	128.56
NET AFTER TAX INCOME	19.11	167.92	101.83	105.65	109.47	113.28	117.10	120.92	124.74	128.56
OPERATING CASH FLOW	118.40	259.57	185.84	182.02	178.20	174.38	170.56	166.75	162.93	159.11
DEBT RETIREMENT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NET CASH FLOW	118.39	259.57	185.84	182.02	178.20	174.38	170.56	166.75	162.93	159.11

TABLE 120 (Continued)
USAF SHALE OIL STUDY
PROFORMA FINANCIAL STATEMENT

ECONOMIC EVALUATION MODEL

RUN NUMBER   1

ALL $ VALUES IN MILLIONS

YEAR	1994	1995	1996	1997	1998	1999
PERCENT CAPACITY	100.00	100.00	100.00	100.00	100.00	100.00
ANNUAL SALES	1352.40	1352.40	1352.40	1352.40	1352.40	1352.40
RAW MATERIAL COST	990.00	990.00	990.00	990.00	990.00	990.00
GROSS MARGIN	362.40	362.40	362.40	362.40	362.40	362.40
OPERATING LABOR	9.21	9.21	9.21	9.21	9.21	9.21
UTILITY COST	16.94	16.94	16.94	16.94	16.94	16.94
MAINTENANCE COST	20.28	20.28	20.28	20.28	20.28	20.28
CAT+CHEM COST	8.92	8.92	8.92	8.92	8.92	8.92
DIRECT MFG EXPENSE	55.35	55.35	55.35	55.35	55.35	55.35
INSURANCE	3.38	3.38	3.38	3.38	3.38	3.38
PROPERTY TAX	6.76	6.76	6.76	6.76	6.76	6.76
INDIRECT MFG EXPENSE	10.14	10.14	10.14	10.14	10.14	10.14
DEPRECIATION	22.91	15.27	7.64			
INTEREST ON W. CAPITAL	9.24	9.24	9.24	9.24	9.24	9.24
TAXABLE INCOME	264.76	272.39	280.03	287.67	287.67	287.67
TAX LOSS DEDUCTION						
ADJ. TAXABLE INCOME	264.76	272.39	280.03	287.67	287.67	287.67
CURRENT TAX	132.38	136.20	140.01	143.83	143.83	143.83
TAX CREDITS	0.	0.	0.	0.	0.	0.
TAXES PAYABLE	132.38	136.20	140.01	143.83	143.83	143.83
NET AFTER TAX INCOME	132.38	136.20	140.01	143.83	143.83	143.83
OPERATING CASH FLOW	155.29	151.47	147.65	143.83	143.83	143.83
DEBT RETIREMENT	0.	0.	0.	0.	0.	0.
RETURN OF WORK. CAP.						93.74
NET CASH FLOW	155.29	151.47	147.65	143.83	143.83	143.83

JP4:15%ROI

TABLE 120 (Continued)

USAF SHALE OIL STUDY

PROFITABILITY SUMMARY

ECONOMIC EVALUATION MODEL
RUN NUMBER   1

ALL $ VALUES IN MILLIONS

PAYBACK PERIOD: YEARS   4.25

AFTER TAX DISCOUNTED
RATE OF RETURN ON EQUITY %   15.00

AFTER TAX DISCOUNTED
RATE OF RETURN ON PROJECT %   15.00

NET PRESENT VALUE @15%: $MM   88.67

N.P.V. INDEX   @15%   0.11

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TABLE 120 (Continued)

BASIS FOR STUDY

PROFITABILITY SUMMARY

ECONOMIC EVALUATION MODEL
RUN NUMBER   1   JP4:15%ROI   PAGE   3 - 1

ALL $ VALUES IN MILLIONS

ECONOMIC PARAMETERS

PROJECT LIFE:YEARS	16
CONSTR PERIOD:YEARS	3
INTEREST RATE: %	0.000
INCOME TAX RATE: %	50.000
PERCENT DEBT	0.00
GENERAL INFLATION FACTOR	1.000
LABOR INFLATION FACTOR	1.000
UTILITY INFLATION FACTOR	1.000
PETRO PRODS INFL FACTOR	1.000
CONST COST INFL FACTOR	1.000
WORKING CAP INFL FACTOR	1.000
INV. TAX CREDIT: % OF CAP	10.000
INV TX CR: MAX % OF TAX	90.000

BASE FEED PRICES

SHALE OIL	30.00

BASE PRODUCT PRICES

JET FUEL $/BBL	42.97
AMMONIA $/ST	120.00
SULFUR $/LT	53.00

BASE WORKING CAPITAL

RAW MATERIALS	68.44
ACCOUNTS RECEIVABLE	0.
ACCOUNTS PAYABLE	0.
NOBLE METALS	0.
LAND	0.
CATALYST&CHEMICALS	25.30
ROYALTY & FEES	0.

TABLE 120 (Continued)

USAF SHALE OIL STUDY

INVESTMENT SUMMARY

ECONOMIC EVALUATION MODEL
RUN NUMBER   1

ALL $ VALUES IN MILLIONS

ESTIMATED CAPITAL INVESTMT   695.00
CONST. INFLATION FACTOR   1.000
INTEREST RATE: %   0.000

	1981	1982	1983	
YEAR				
% COMPLETION FOR YR	25.00	50.00	25.00	
PLANT INVESTMENT				
. CAPITAL INVESTMENT	173.75	347.50	173.75	
. CONST. COST INFLATION	-0.00	-0.00	-0.00	
. CONST. COST FOR YEAR	173.75	347.50	173.75	
TOTAL DEPRECIABLE INV.				695.00
OTHER INVESTMENT				
. LAND				0.
. INITIAL CAT.&CHEM.				25.30
. ROYALTY & FEES				0.
SUBTOTAL				25.30
WORKING CAPITAL				
. RAW MATERIALS				68.44
. ACCOUNTS RECEIVABLE				0.
. ACCOUNTS PAYABLE				0.
. NOBLE METALS				0.
. INT ON CONSTRUCTION	0.	0.	0.	
TOTAL CONST INTEREST				0.
SUBTOTAL				68.44
TOTAL INVESTMENT				788.74

ECONOMIC EVALUATION MODEL

RUN NUMBER   1
ALL $ VALUES IN MILLIONS

TABLE 120 (Continued)

USAF SHALE OIL STUDY
PRODUCTION SUMMARY

YEAR	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
	50.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
PRODUCTION,MM UNITS/YEAR										
JET FUEL (BBLS)	15.58	31.16	31.16	31.16	31.16	31.16	31.16	31.16	31.16	31.16
AMMONIA (ST)	0.05	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10	0.10
SULFUR (LT)	0.01	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03
PRODUCT PRICES: $/UNIT										
JET FUEL $/BBL	42.97	42.97	42.97	42.97	42.97	42.97	42.97	42.97	42.97	42.97
AMMONIA $/ST	120.00	120.00	120.00	120.00	120.00	120.00	120.00	120.00	120.00	120.00
SULFUR $/LT	53.00	53.00	53.00	53.00	53.00	53.00	53.00	53.00	53.00	53.00
PRODUCT VALUES $MM/YEAR										
JET FUEL	669.41	1338.82	1338.82	1338.82	1338.82	1338.82	1338.82	1338.82	1338.82	1338.82
AMMONIA	2.15	4.30	4.30	4.30	4.30	4.30	4.30	4.30	4.30	4.30
SULFUR	0.64	1.28	1.28	1.28	1.28	1.28	1.28	1.28	1.28	1.28

TABLE 120 (Continued)

USAF SHALE OIL STUDY
PRODUCTION SUMMARY

ECONOMIC EVALUATION MODEL

RUN NUMBER 1
ALL $ VALUES IN MILLIONS

YEAR	1994	1995	1996	1997	1998	1999
	100.00	100.00	100.00	100.00	100.00	100.00
PRODUCTION,MM UNITS/YEAR						
JET FUEL (BBLS)	31.16	31.16	31.16	31.16	31.16	31.16
AMMONIA (ST)	0.10	0.10	0.10	0.10	0.10	0.10
SULFUR (LT)	0.03	0.03	0.03	0.03	0.03	0.03
PRODUCT PRICES: $/UNIT						
JET FUEL $/BBL	42.97	42.97	42.97	42.97	42.97	42.97
AMMONIA $/ST	120.00	120.00	120.00	120.00	120.00	120.00
SULFUR $/LT	53.00	53.00	53.00	53.00	53.00	53.00
PRODUCT VALUES $MM/YEAR						
JET FUEL	1338.82	1338.82	1338.82	1338.82	1338.82	1338.82
AMMONIA	4.30	4.30	4.30	4.30	4.30	4.30
SULFUR	1.28	1.28	1.28	1.28	1.28	1.28

TABLE 121

USAF SHALE OIL STUDY
PROFORMA FINANCIAL STATEMENT

JP8:15XROI

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ECONOMIC EVALUATION MODEL

RUN NUMBER 2

ALL $ VALUES IN MILLIONS

YEAR	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
PERCENT CAPACITY	50.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
ANNUAL SALES	684.40	1368.81	1368.81	1368.81	1368.81	1368.81	1368.81	1368.81	1368.81	1368.81
RAW MATERIAL COST	495.00	990.00	990.00	990.00	990.00	990.00	990.00	990.00	990.00	990.00
GROSS MARGIN	189.41	378.81	378.81	378.81	378.81	378.81	378.81	378.81	378.81	378.81
OPERATING LABOR	10.59	10.59	10.59	10.59	10.59	10.59	10.59	10.59	10.59	10.59
UTILITY COST	8.21	16.43	16.43	16.43	16.43	16.43	16.43	16.43	16.43	16.43
MAINTENANCE COST	21.18	21.18	21.18	21.18	21.18	21.18	21.18	21.18	21.18	21.18
CAT-CHEM COST	4.56	9.10	9.10	9.10	9.10	9.10	9.10	9.10	9.10	9.10
DIRECT MFG EXPENSE	44.54	57.30	57.30	57.30	57.30	57.30	57.30	57.30	57.30	57.30
INSURANCE	3.53	3.53	3.53	3.53	3.53	3.53	3.53	3.53	3.53	3.53
PROPERTY TAX	7.06	7.06	7.06	7.06	7.06	7.06	7.06	7.06	7.06	7.06
INDIRECT MFG EXPENSE	10.59	10.59	10.59	10.59	10.59	10.59	10.59	10.59	10.59	10.59
DEPRECIATION	103.71	95.74	87.76	79.78	71.80	63.82	55.85	47.87	39.89	31.91
INTEREST ON W. CAPITAL	9.24	9.24	9.24	9.24	9.24	9.24	9.24	9.24	9.24	9.24
TAXABLE INCOME	21.32	205.94	213.92	221.90	229.88	237.85	245.83	253.81	261.79	269.77
TAX LOSS DEDUCTION										
ADJ. TAXABLE INCOME	21.32	205.94	213.92	221.90	229.88	237.85	245.83	253.81	261.79	269.77
CURRENT TAX	10.66	102.97	106.96	110.95	114.94	118.93	122.92	126.91	130.89	134.88
TAX CREDITS	9.60	62.99	0.	0.	0.	0.	0.	0.	0.	0.
TAXES PAYABLE	1.06	39.98	106.96	110.95	114.94	118.93	122.92	126.91	130.89	134.88
NET AFTER TAX INCOME	20.26	165.96	106.96	110.95	114.94	118.93	122.92	126.91	130.89	134.88
OPERATING CASH FLOW	123.97	261.70	194.72	190.73	186.74	182.75	178.76	174.77	170.78	166.80
DEBT RETIREMENT	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
NET CASH FLOW	123.97	261.69	194.72	190.73	186.74	182.75	178.76	174.77	170.78	166.79

TABLE 121 (Continued)

USAF SHALE OIL STUDY
PROFORMA FINANCIAL STATEMENT

ECONOMIC EVALUATION MODEL

RUN NUMBER 2   JP8:15%ROI

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ALL $ VALUES IN MILLIONS

YEAR	1994	1995	1996	1997	1998	1999
PERCENT CAPACITY	100.00	100.00	100.00	100.00	100.00	100.00
ANNUAL SALES	1368.81	1368.81	1368.81	1368.81	1368.81	1368.81
RAW MATERIAL COST	990.00	990.00	990.00	990.00	990.00	990.00
GROSS MARGIN	378.81	378.81	378.81	378.81	378.81	378.81
OPERATING LABOR	10.59	10.59	10.59	10.59	10.59	10.59
UTILITY COST	16.43	16.43	16.43	16.43	16.43	16.43
MAINTENANCE COST	21.18	21.18	21.18	21.18	21.18	21.18
CAT+CHEM COST	9.10	9.10	9.10	9.10	9.10	9.10
DIRECT MFG EXPENSE	57.30	57.30	57.30	57.30	57.30	57.30
INSURANCE	3.53	3.53	3.53	3.53	3.53	3.53
PROPERTY TAX	7.06	7.06	7.06	7.06	7.06	7.06
INDIRECT MFG EXPENSE	10.59	10.59	10.59	10.59	10.59	10.59
DEPRECIATION	23.93	15.96	7.98			
INTEREST ON W. CAPITAL	9.24	9.24	9.24	9.24	9.24	9.24
TAXABLE INCOME	277.74	285.72	293.70	301.68	301.68	301.68
TAX LOSS DEDUCTION						
ADJ. TAXABLE INCOME	277.74	285.72	293.70	301.68	301.68	301.68
CURRENT TAX	138.87	142.86	146.85	150.84	150.84	150.84
TAX CREDITS	0.	0.	0.	0.	0.	0.
TAXES PAYABLE	138.87	142.86	146.85	150.84	150.84	150.84
NET AFTER TAX INCOME	138.87	142.86	146.85	150.84	150.84	150.84
OPERATING CASH FLOW	162.81	158.82	154.83	150.84	150.84	150.84
DEBT RETIREMENT	0.	0.	0.	0.	0.	0.
RETURN OF WORK. CAP.						94.20
NET CASH FLOW	162.81	158.82	154.83	150.84	150.84	150.84

TABLE 121 (Continued)

USAF SHALE OIL STUDY

PROFITABILITY SUMMARY

JP8:15%ROI

PAGE 2 - 1

ECONOMIC EVALUATION MODEL
RUN NUMBER   2

ALL $ VALUES IN MILLIONS

PAYBACK PERIOD: YEARS   4.27

AFTER TAX DISCOUNTED
RATE OF RETURN ON EQUITY %   15.00

AFTER TAX DISCOUNTED
RATE OF RETURN ON PROJECT %   15.00

NET PRESENT VALUE @15%: $MM   92.15

N.P.V. INDEX   @15%   0.11

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TABLE 121 (Continued)

BASIS FOR STUDY

PROFITABILITY SUMMARY

JP8:15%ROI

ECONOMIC EVALUATION MODEL
RUN NUMBER   2

ALL $ VALUES IN MILLIONS

ECONOMIC PARAMETERS

PROJECT LIFE:YEARS	16
CONSTR PERIOD:YEARS	3
INTEREST RATE: %	0.000
INCOME TAX RATE: %	50.000
PERCENT DEBT	0.00
GENERAL INFLATION FACTOR	1.000
LABOR INFLATION FACTOR	1.000
UTILITY INFLATION FACTOR	1.000
PETRO PRODS INFL FACTOR	1.000
CONST COST INFL FACTOR	1.000
WORKING CAP INFL FACTOR	1.000
INV. TAX CREDIT: % OF CAP	10.00
INV TX CR: MAX % OF TAX	90.000

BASE FEED PRICES

SHALE OIL	30.00

BASE PRODUCT PRICES

JET FUEL $/BBL	45.40
AMMONIA $/ST	120.00
SULFUR $/LT	53.00

BASE WORKING CAPITAL

RAW MATERIALS	67.35
ACCUONTS RECEIVABLE	0.
ACCOUNTS PAYABLE	0.
NOBLE METALS	0.
LAND	0.
CATALYST&CHEMICALS	26.85
ROYALTY & FEES	0.

TABLE 121 (Continued)

USAF SHALE OIL STUDY

INVESTMENT SUMMARY

JP8:15%ROI

ECONOMIC EVALUATION MODEL
RUN NUMBER 2

ALL $ VALUES IN MILLIONS

ESTIMATED CAPITAL INVESTMT	726.00
CONST. INFLATION FACTOR	1.000
INTEREST RATE: %	0.000

YEAR	1981	1982	1983
% COMPLETION FOR YR	25.00	50.00	25.00

PLANT INVESTMENT

	1981	1982	1983
. CAPITAL INVESTMENT	181.50	363.00	181.50
. CONST. COST INFLATION	-0.00	-0.00	-0.00
. CONST. COST FOR YEAR	181.50	363.00	181.50

| TOTAL DEPRECIABLE INV. | 726.00 |

OTHER INVESTMENT

. LAND	0.
. INITIAL CAT.&CHEM.	26.85
. ROYALTY & FEES	0.
SUBTOTAL	26.85

WORKING CAPITAL

. RAW MATERIALS			67.35
. ACCOUNTS RECEIVABLE			0.
. ACCOUNTS PAYABLE			0.
. NOBLE METALS			0.
. INT ON CONSTRUCTION			0.
. TOTAL CONST INTEREST	0.	0.	0.
SUBTOTAL			67.35

| TOTAL INVESTMENT | 820.20 |

ECONOMIC EVALUATION MODEL

JP8:15BROI

TABLE 121 (Continued)

USAF SHALE OIL STUDY
PRODUCTION SUMMARY

PAGE 5 - 1

RUN NUMBER 2
ALL $ VALUES IN MILLIONS

	1984	1985	1986	1987	1988	1989	1990	1991	1992	1993
YEAR	50.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
PRODUCTION,MM UNITS/YEAR										
JET FUEL (BBLS)	14.93	29.87	29.87	29.87	29.87	29.87	29.87	29.87	29.87	29.87
AMMONIA (ST)	0.05	0.09	0.09	0.09	0.09	0.09	0.09	0.09	0.09	0.09
SULFUR (LT)	0.01	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03	0.03
PRODUCT PRICES: $/UNIT										
JET FUEL $/BBL	45.40	45.40	45.40	45.40	45.40	45.40	45.40	45.40	45.40	45.40
AMMONIA $/ST	120.00	120.00	120.00	120.00	120.00	120.00	120.00	120.00	120.00	120.00
SULFUR $/LT	53.00	53.00	53.00	53.00	53.00	53.00	53.00	53.00	53.00	53.00
PRODUCT VALUES $MM/YEAR										
JET FUEL	678.03	1356.06	1356.06	1356.06	1356.06	1356.06	1356.06	1356.06	1356.06	1356.06
AMMONIA	2.11	4.22	4.22	4.22	4.22	4.22	4.22	4.22	4.22	4.22
SULFUR	0.68	1.36	1.36	1.36	1.36	1.36	1.36	1.36	1.36	1.36

TABLE 121 (Continued)

ECONOMIC EVALUATION MODEL JP8:15XROI

RUN NUMBER 2
ALL $ VALUES IN MILLIONS

USAF SHALE OIL STUDY
PRODUCTION SUMMARY

YEAR	1994	1995	1996	1997	1998	1999
	100.00	100.00	100.00	100.00	100.00	100.00
PRODUCTION,MM UNITS/YEAR						
JET FUEL (BBLS)	29.87	29.87	29.87	29.87	29.87	29.87
AMMONIA (ST)	0.09	0.09	0.09	0.09	0.09	0.09
SULFUR (LT)	0.03	0.03	0.03	0.03	0.03	0.03
PRODUCT PRICES: $/UNIT						
JET FUEL $/BBL	45.40	45.40	45.40	45.40	45.40	45.40
AMMONIA $/ST	120.00	120.00	120.00	120.00	120.00	120.00
SULFUR $/LT	53.00	53.00	53.00	53.00	53.00	53.00
PRODUCT VALUES $MM/YEAR						
JET FUEL	1356.06	1356.06	1356.06	1356.06	1356.06	1356.06
AMMONIA	4.22	4.22	4.22	4.22	4.22	4.22
SULFUR	1.36	1.36	1.36	1.36	1.36	1.36

APPENDIX A

Fill Procedure

1. Plant power off. Make sure utilities are on- air, N2 purge, water plant slop tank should be empty and connected to plant drain system.

2. Vent plant pressure at charge tank V-1. Close vent valve when done. Open main drain exit valve. Open DV-5 for 1 minute then close. Close main drain exit valve.

3. PC-7 full CCW then 3 turns CW. PC-8 full CCW, then 5 turns CW.

4. Pressure transfer vessel with 35 lbs N2. Hook transfer line to bottom of transfer vessel and top of charge tank.

5. Open hand valve on top of charge tank at transfer line.

6. Close vent valve on top of charge tank.

7. Adjust PC-6 to full CW.

8. Plant power on select drain mode.

9. Enable plant power.

10. Open valve on bottom of transfer vessel to allow liquid into plant.

11. Watch charge tank pressure gauge till 35 lbs.

12. Make sure main drain exit valve to slop tank is open. Slowly open DV-5 to allow liquid to fill back end of plant. Watch for liquid coming out of DV-5. Then close DV-5 and valve at bottom of transfer vessel.

13. Vent pressure on charge tank to atmosphere by opening vent valve on top of charge tank slowly. Leave open.

14. Add enough liquid till it starts blowing over from the top of the plant to the slop can.

15. Move transfer line to top of transfer vessel and blow down line into charge tank by opening transfer valve at top of transfer vessel.

16 Close transfer line valve at top of charge tank.

17. Remove transfer line from plant and depressure.

18. If solvent has been put into plant go to step 21.

19. Close vent valve at top of charge tank.

20. Open PC-6 full CCW. Set sparge rate and adjust PC-6 to desired pressure.

21. Run --fill/drain switch to run position.

22. Plant now ready to run.

Drain Procedure

1. Plant power off. Make sure utilities are on-air, N2, water. Plant slop tank should be empty and connected to plant drain system. Vent plant pressure at charge tank V-1. Close vent valve when done. Open main drain exit valve. Open DV-5 for 1 minute then close. Close main drain exit valve.

2. PC-7 full CCW. Then 3 turns CW. PC-8 full CCW. Then 5 turns CW.

3. Pressure transfer vessel with 35 lbs N2.

4. Hook transfer line to top of transfer tank and to top of charge tank V-1.

5. Plant power. On select drain mode with switch. Enable plant power.

6. Make sure vent valve on top of plant is closed. Adjust PC-6 full CW.

7. Open transfer line on top of charge tank V-1. Slowly open transfer line on top of transfer vessel and pressure charge tank to 35 lbs.

8. Open main drain exit valve.

9. Open DV-7 and allow plant charge tank to drain. Close DV-7 when draining complete.

10. Open DV-5 and allow plant to drain. Close DV-5 when draining is complete.

11. PC-7 full CCW. Slowly open DV-2 and allow E-4 to drain. Close DV-2 when done. PC-7 3 turns CW.

12. Open and drain the following valves in order: DV-3, DV-1, DV-4, DV-6, DV-7. Close valve when done draining.

13. Close valve on transfer line at top of plant and at top of transfer vessel. Remove line and depressure.

14. Slowly adjust PC-6 CCW until pressure on charge tank is 10 lbs. Open DV-5 and allow plant to depressure. Leave DV-5 open.

15. Slowly PC-8 full CCW allow turbine meter leg to drain. Watch electronics cabinet and never allow turbine meter indicator to exceed 100. Close DV-5 when complete.

16. PC-8 - 5 turns CW. Close main drain exit valve.

17. Plant is now drained.

Flush Procedure

1. Warm up hot box to 100 Deg. F to 110 Deg. F if not already warm.

2. Drain plant according to drain procedure.

3. Remove then pump and turbine meter filters. Install clean ones.

4. Remove fouling cell and install pipe bypass tube.

5. Fill plant with flush solvent according to fill procedure.

6. Run the plant at the following conditions:

    A. Pump discharge = 150 Lbs. <PC-7>
    B. Plant pressure = 75 Lbs. <PC-7>
    C. Flow rate = 600 ML/Min.
    D. Fluid Temp = 0
    E. Period 2 time = 2 Hr.
    F. Report rate   = 1 Min.
    G. Vent valve on charge tank is open
    H. No sparge rate is needed.
    I. Cooling water temp = 110 Deg. F.
    J   Cooling water flow rate = 30% <1 GPM>
    K. Purge gas flow rate = 2 on rotameter

UOP PROCESS/MONIREX

LABORATORY FOULING MONITOR

DEFINITIONS
============

Test	A collection of periods that start with period 1 and end with the end period.
Period	A time slice in which parameters are executed.
End period	The last period in a test. The end period is never executed, period 1 is executed in its place.
Period 1	The null period. If a test is not running, period 1 is the active period. The parameters for period 1 are not changeabl
Active period	The period being executed.

TYPE AHEAD AND LINE EDITING
==============================

Type-ahead allows the entering of up to 63 characters (without an echo) even when there is no current request for input.

Line editing assures the receiving of relatively 'clean' input from the terminal. There is no direct hardware data link between the keyboard and the printer; rather, each keyboard input to the computer is echoed back, by software to the console for printing. This technique assures the intended character has been entered without error. The entered (input) characters and the echoed (output) characters are stored in separate buffers until a carriage return (RETURN) is typed or a maximum of 132 characters have been entered. In either case, the input (line editing) buffer content is forwarded to its intended destination via one or more of the system generated read commands. Prior to line termination, the buffer content can be revised or deleted by using special non-printable 'editing' characters. Theses 'editing' characters are not normally stored in the line editing buffer but rather provide control over the buffer contents. The following characters are used the edit the input of the line-edited buffer. Control characters are entered by holding down the control key (CTRL) while the character is typed.

RUBOUT	Deletes the last character from the line buffer and echoes the deleted character to the display. The bell sounds is an attempt is made to RUBOUT past the begining of the logical line.
control-C	A special character not to be used.
control-O	Kills the output. All output after the control-O is issued is lost until a subsequent control-O is entered.

```
control-P        A special character not to be used.

control-Q        Resumes terminal output (see control-S).

control-S        Suspends the display of output.  It does not affect
                 alarm messages.  The data cannot be displayed
                 until a control-Q is issued.  CAUTION: Prolonged
                 usage of control-S will cause the system to degrade
                 such as inhibiting advancement to the next period.

control-R        Causes the display of the current contents of the
                 line editing buffer.  Control-R issues a carraige
                 return and line feed and thens echoes a copy of all
                 characters for that line.  Control-R is usefull
                 in generating a clean copy of the current line,
                 minus all the characters that have been deleted
                 using RUBOUT.

control-X        Deletes the entire contents of the line edit buffer.
                 A number sign (#) followed by a carraige return and
                 line feed is displayed.  Control-X is useful for
                 correcting errors that are inconvient to correct
                 using RUBOUT.

control-Z        Causes an end-of-file to be input to the system.
                 There is no change to the display.  The contents
                 of the line editing buffer are deleted.


POWER UP
========
The date and time are solicited upon power up to the computer.
Four solicits are made:
MONTH    The month number (1=January, 2=February,... 12=December).
DAY      The day of the month.
HOUR     The hour of the day (24 hour clock).
MINUTE   The minute of the hour.


COMMAND LANGUAGE
================
The command language provides the operator interface to the computer.
All comands are terminated by a carriage return except for a control-Z.
Commands are solicited with a 'greater than' (>) symbol.


Command Summary
----------------
SCHEDULE         Enters parameters into the test table.
TABLE            Lists the test table.
PERIOD           Edits a period.
RUN              Initiates the test.
SKIP             Terminates the current activer period.
SHUTDOWN         Shuts down the plant.
```

SCHEDULE - Scheduling a Test

The SCHEDULE command allows the entry of period parameters into the
test table. The parameters for each period are prompted with a star (*)
character. The syntax of the SCHEDULE command is:

           SCHEDULE

A test title of up to 78 characters is solicited. There are six parameters
which are solicited for each period:

CALIBRATION   If the response to this prompt is a "Y" then the period
                 is a calibration period and no other parameters are
                 solicited; otherwise, the remaining parameters are
                 solicited.

TIME   The duration of the period.

FLOW   The flow rate setpoint for the fluid at the probe.

FL TEMP   The fluid temperature setpoint for the fluid leaving
                 the H-2 heater.

VOLTAGE   The probe voltage setpoint.

REPORT   The report interval.

For each parameter (except CALIBRATION), a new value may be entered
or the listed default used. The default is taken from the previous
period's corresponding parameter. To use the default, a slash (/)
is entered and then a return. The prompting is terminated by the
maximum number of periods or a control-Z. A control-Z marks the
period as the end period. Parameters entered for that period (if any),
are ignored.

TABLE - Viewing the Test Table

The TABLE command list the test table. The table command discards
report output during its execution. The syntax of the TABLE
command is:

           TABLE

PERIOD - Editing a Period

The PERIOD command permits a period to be edited. The period number is
solicited with the end period being the default. The remaining parameters
are solicited in the same as the SCHEDULE command except only one period is

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changed and the default values are taken from the period being edited instead of from the previous period. An exceptin occures for the end period, its defaults are taken from the previous period's. Only the following period types may be changed:
1. an active period
2. the end period (the end period is bumped)
3. an unexecuted period but is before the end period
If a control-Z is entered for an active period, the new parameters are ignored, the period reverts back to its old parameters, and the period is not restarted. If the period is not an active period and a control-Z is entered, the period is marked as an end period. If an active period is edited, the period will be restarted. This may make it necessary to change the time parameter. The period command discards report output during its execution. The syntax of the PERIOD command is:

        PERIOD

RUN - Running a Test

The RUN command initates a test. The test always starts with period 2 and ends with the end period; however, the end period is never executed (period 1 is executed in its place). If the RUN command is issued during the test, period 2 will become the active period. The syntax of the RUN command is:

        RUN

SKIP - Skipping a Period

The SKIP command causes the current active period to terminate. The next period will be run if it is not the end period; otherwise, period 1 will be executed. The syntax of the SKIP command is:

        SKIP

SHUTDOWN - Shutting Down the Plant

The SHUTDOWN command shuts the plant down. The following actions are performed:
1. pumps are turned off
2. control valve is closed
3. heaters are turned off
4. probe voltage is set to zero volts
5. period 1 is the active period
Shutting down an already shut down plant has no effect. The syntax of the SHUTDOWN command is:

        SHUTDOWN
        -- or --
        SHUT DOWN

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LIMITS

Test Table Parameter Limits

Parameter	Minimum	Maximum	Resolution	Unit	Notes
PERIOD	2	50			
TIME	0.008	540.000	0.017	hour	a,b
FLOW	10.	850.	1.	ml/min	a,c
FL TEMP	1.0	315.0	0.5	C	a,d
VOLTAGE	0.001	5.000	0.002	V	
REPORT	0.008	540.000	0.017	min	a,e

a. A value greater than or equal to zero and less than the minimum is permissible.
b. Time less than the minimum skips the period.
c. Flow less than the minimum will not turn the pump on ·
d. Fluid temperature less than the minimum will no turn the heaters on regardless of the actual temperature.
e. Report less than the minimum inhibits reporting.

Report Parameter Limits

Parameter	Minimum	Maximum	Resolution	Unit	Notes
Elapsed time	0.000	540.000	0.017	hour	a
Probe temp			0.5	C	
Probe voltage	0	5.000	0.002	V	
Probe current	0	25.000	0.006	A	
Other temps			1.	C	

a. Elapsed time is incremented internally in intergral minutes.

FAULTS

Alarm Messages

The following are alarm messages and are listed in priority order (the first is the highest priority):

PLANT POWER
        The plant power is off.

RESERVOIR LEVEL
        The fluid level in the reservoir is low.

WATER FLOW
        The cooling water flow is too low.

PLANT PRESSURE
> The pump discharge pressure is too high.

RESERVOIR TEMPERATURE
> The fluid temperature in the reservoir is too high.

HOT BOX TEMPERATURE
> The temperature of the hot box is too high.

HEATER TEMPERATURE
> The block temperature of one or both heaters is too high.

FLOW CONTROLLER
> The flow controller is unable to cope with hardware conditions. The probe fluid flow rate has not responded as expected to the control valve position commands.

PROBE VOLTAGE CONTROLLER
> The probe voltage controller is unable to cope with hardware conditions. Sources of the failure are:
> 1. The probe fluid flow rate was less than 45 ml/min upon entry into the calibration period.
> 2. The probe voltage did not respond properly within the timeout for the calibration period.
> 3. The difference between the setpoint and the actual voltage is too great.
> 4. The controller is calling for a voltage the hardware cannot deliver.

H2 CONTROLLER
> The H-2 heater temperature controller is unable to cope with hardware conditions. The fluid or block temperature for
> H-2 is above 325 C.

H1 CONTROLLER
> The H-1 heater temperature controller is unable to cope with hardware conditions. The fluid temperature is above 325 C.

RUN SWITCH
> The RUN/FILL switch has been changed to FILL during a test.

APPENDIX C

OFFSITE EQUIPMENT

A summary of the offsite equipment to be included in the refinery is as follows:

A. **Fresh Water System**

   Distribution system for boiler feed, pump cooling, laboratory, and personnel use.

B. **Closed Water Cooling System**

   The circulating and cooling systems for trim cooling are to be provided in the process units (refinery is to be maximum air cooled).

C. **Fire Fighting Facilities**

   Supply and distribution system using fresh water for the fire grid is to be provided. This will include hydrants as well as portable and mobile fire-fighting equipment.

D. **Refinery Fuel System**

   Provide fuel gas and fuel oil gathering and distribution systems.

E. **Electric Power System**

   Includes substations and distribution system with emergency supply for instruments.

F. **Compressed Air System**

   Instrument and plant air system.

G. **Waste Disposal Systems**

   Includes blowdown system and flare, oil sewer system including API separator, secondary treatment, sanitary waste system, storm system, waste water stripper and caustic neutralization.

H. **Shale Oil and Product Storage**

   Provide minimum of 21 days of crude storage and 14 days of product storage. Include also intermediate tankage between process units. Tankage, includes accessories, painting, insulation and heaters where required, remote gauging and remote operators on valves.

I. _Transfer and Product Pumps_

J. _Yard Oil Piping_

Includes piping from crude tanks to process area and piping to product tanks, slop lines to storage, gas blanketing lines, inter-unit piping and yard piping between tankage.

K. _Inert Gas System_

L. _Buildings_

Control house, laboratory, office, change house,pump house if required, maintenance shops and fire house.

M. _Fencing and Gates_

N. _Roads and Parking_

O. _Steam Generation and Distribution System_

System includes makeup and letdown stations to satisfy various pressure levels, spare boiler capacity for servicing and feed-water treatment plant for providing acceptable boiler feed water.

P. _Communications System_

APPENDIX D

ESTIMATED ERECTED COST BASIS

The estimated erected cost presented in this study reflects a current US Gulf Cost battery limits price corrected to Midwest, USA location. It is comprised of a materials and labor (M α L) estimate and a design, engineering and contractor's fee, overheads and expenses (DE α CE) allowance.

The material and labor estimates have been derived by scaling detailed estimates prepared for similar units on the basis of US Gulf Coast erection to UOP Standards and Specifications. The material and labor estimates are intended to include all direct material and labor, indirect field costs and labor benefits that are associated with the erection of the battery limits process equipment, including the following specific equipment, categories and services as and when required.

Heaters	Compressor shelter
Vessels and internals	Control house
Heat exchange equipment	Catalyst handling equipment
Pumps	Sundry construction equipment
Drivers	Temporary field office,
Compressors	warehouse, change house, etc.
Piping	Field testing
Instruments	Expendable tools
Electrical	Clerical costs associated with
Insulation	construction
Structural Steelwork	Final cleaning
Fireproofing	Miscellaneous field costs
Paving and concrete work	Fringe benefits

An allowance for design, engineering and contractor's fees, overheads and expenses, primarily based on past UOP experience, has been added to the total material and labor estimate in order to reach an overall erected cost estimate for the battery limits plant. The figure shown for this DE and CE allowance is for orientation economic purposes only and is intended to cover the following charges:

UOP:	Basic process and engineering design specifications and drawings (Schedule A package) including review of contractor's detailed design of specified equipment items.
Contractor:	Detailed engineering Purchasing, expediting and inspection Construction tools and equipment rental Contractor's field and home office expenses Erection supervision Contractor's fees

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Items not included in the estimated UOP investment cost (battery limits) are as follows, unless otherwide specified as included:

- Docks, marine terminals, or jetties.
- Access roads to site.
- Home office administration building.
- Employees housing, worker's barracks, canteens, recreation facilities, etc.
- Overtime pay during construction.
- Know-how fees and royalties on licensed processes.
- Owner's expenses in developing the project.
- Local permits, taxes and fees, or specific costs of doing business in the area.
- Items concerned with export shipment, such as ocean freight, marine insurance, import taxes, customs, etc.
- Operating capital and investment in goods and prices.
- Escalation on materials and labor due to price fluctuation or economic conditions.
- Contingencies.
- Cost of start-up including testing, manpower, utilities, operating manuals and training programs.
- Spare parts, special tools or maintenance equipment.
- Catalyst, chemicals and raw materials including initial fill.
- Customer or national standards, codes.
- Special pollution or noise control facilities.
- Electrical main substations.
- Power generation.
- Water or hydrocarbon pipelines.
- For process unit revamps or individual process units: additions or extensions to utilities systems or offsites.
- Laboratory supplies.
- Special communications or computer systems.

The following assumptions are normally made regarding economic conditions at the time the job is bid:

- There will be an adequate supply of skilled labor available for construction.
- There will be competitive bidding by contractors.
- The plant will be constructed in the United States of America.
- There is no lost time due to climatic conditions.
- Material and labor prices are based on the data of the estimate.

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